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Radian Corp, Austin, Tex.

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(Volume II. Technical Discussion)

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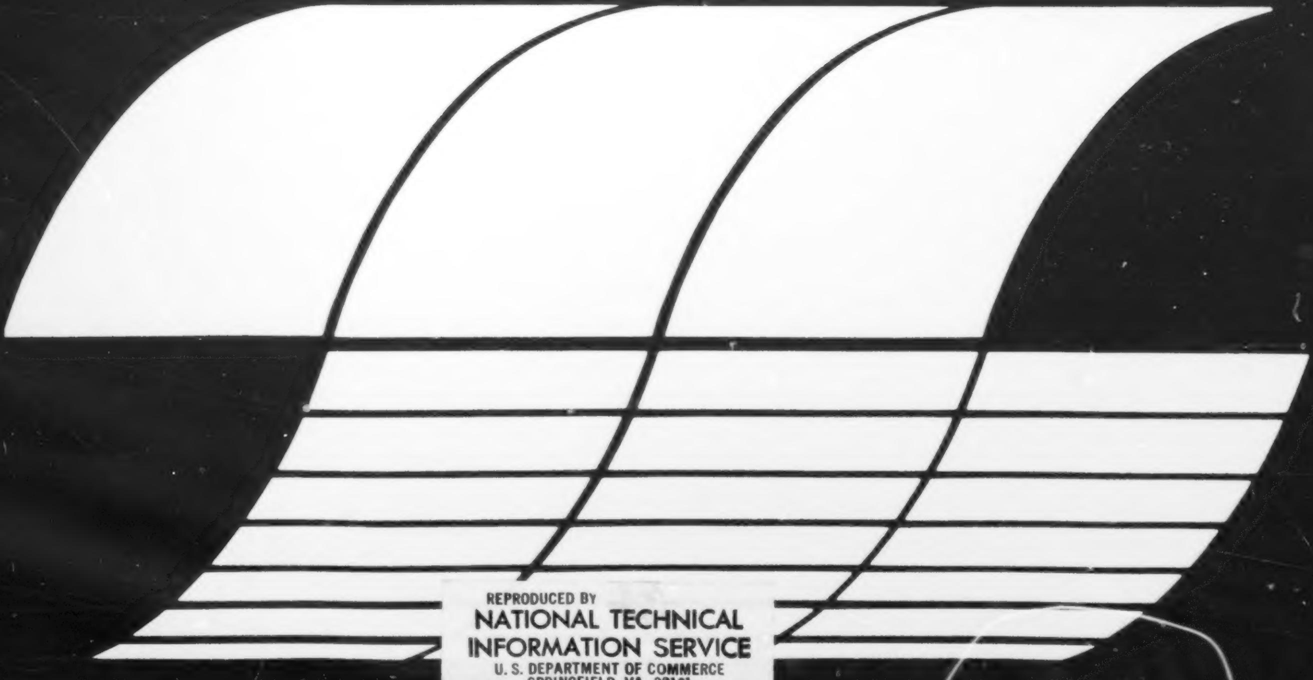

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16. ABSTRACT The report gives results of one task in a comprehensive program to review the New Source Performance Standards (NSPS) for SO2 emissions from coal-fired steam-electric generating plants. The results compare two alternative standards to the existing NSPS (1.2 lb SO2/million Btu of heat input): (1) 0.5 lb SO2/million Btu of heat input, allowing credit (as does the existing NSPS) for physical coal cleaning or use of low sulfur coal; and (2) 90% removal of SO2 from stack gases, regardless of original coal sulfur content. The comparisons are in terms of their effect on the quality and quantity of power plant wastewater effluents and on the amount of plant water consumption. Potential effects of SO2 control system effluents on the environment are evaluated, and alternative treatment processes are discussed. A total of 108 plant systems were discussed, including combinations of three NSPS, five flue gas desulfurization (FGD) systems, five coal types, four plant sizes, and sulfur removal by coal cleaning. Volumes and quality of wastewater streams varied very little from one alternative NSPS to another; all streams can be treated adequately using commercially available technologies. However, the alternative standards increase total water consumption 8-11%, depending on the FGD process used. Physi-coal cleaning plus lime/limestone scrubbing increases total water consumed 8-12%.

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a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Combustion	Pollution Control	13B	
Sulfur Dioxide	Steam-Electric	Stationary Sources	07B	
Flue Gases	Power Generation		21B	10A
Desulfurization	Calcium Oxides		07A, 07D	
Water Pollution	Limestone			
Coal	Wastewater		08G, 21D	
Waste Treatment				

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(Volume II. Technical Discussion)

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LIST OF ABBREVIATIONS

acf	actual cubic foot
BOD	Biochemical Oxygen Demand
Btu	British thermal unit
^oC	degrees Centigrade
cf	cubic foot
cm^3	cubic centimeter
COD	Chemical Oxygen Demand
EPA	Environmental Protection Agency
^oF	degrees Fahrenheit
FGD	Flue Gas Desulfurization
g	gram
gpm	gallons per minute
gal	gallon
J	Joule
Kcal	kilocalories
kg	kilogram
l	liter
L/G	liquid to gas ratio
lb	pound
m^3	cubic meters
mg	milligrams
MJ	megajoule
MM Btu	million Btu
MPa	megapascals
MW	megawatt
Nm^3	normal cubic meter
NSPS	New Source Performance Standard
OAQPS	Office of Air Quality Planning and Standards
Pa	Pascal
ppm	parts per million
psi	pounds per square inch

s	second
scf	standard cubic foot
ss	suspended solids
TDS	Total Dissolved Solids
μg	microgram

METRIC CONVERSION FACTORS

To Convert From	To	Multiply By
Btu/lb	J/kg	2.324×10^{3}
gpm	m^{3}/s	6.309×10^{-5}
gal	m^{3}	3.785×10^{-3}
gal/1000 scf	m^{3}/Nm^{3}	.000135
lb/min	kg/s	7.567×10^{-3}
lb/10^{6} Btu	µg/J	0.4303
psi	MPa	6.9×10^{-3}

1.0 INTRODUCTION

1.1 Objective and Background

The objective of this study is to define and assess the effects of alternative SO_x control systems on the water consumption and wastewaters from coal-fired power plants. The approach used was to calculate the water requirements for model power plant water systems and model SO_x control systems provided by EPA. Effluent streams from uncontrolled coal-fired power plants and each of the processes involved in the SO_x control strategies were characterized. A comparison of the calculated water requirements and characteristic wastewaters of the power plant and the SO_x control systems was made to assess the impact on receiving waters.

The processes comprising the alternative SO_x control systems examined in this study include:

- Lime Wet Scrubbing
- Limestone Wet Scrubbing
- Wellman-Lord Sulfite Scrubbing
- Magnesium Slurry Absorption
- Double Alkali Wet Scrubbing
- Physical Coal Cleaning
- SO_2 Conversion

This study is one task in a comprehensive program by the Office of Air Quality Planning and Standards to review the New Source Performance Standards for SO_2 emissions from coal-fired steam generating plants. The comprehensive impacts of two alternative revised standards, and the existing NSPS are being examined. The existing NSPS allows an emission rate of 0.52 $\mu g\ SO_2/J$

(1.2 lb SO_2/MM Btu) of heat input. One alternative standard requires 0.22 μg SO_2/J (0.5 lb SO_2/MM Btu) of heat input. This standard has the same form as the existing NSPS and thus allows a credit for physical coal cleaning or use of low sulfur coal. The second alternative standard requires 90% removal of SO_2 from stack gases, regardless of original sulfur content in the coal. 108 model plant system permutations were defined for evaluation. These model systems are discussed in detail in Section 4. Generally, these model systems allow an analysis of the impacts of the three alternative NSPS for SO_2, accommodating the variables of; type of flue gas desulfurization (FGD) system (5 types), sulfur content of the coal, size of the steam generator (25,100,500 and 1000 Mw), and degree of coal cleaning.

An examination of the model plant water consumption calculations indicates that the application of the SO_x control systems to coal-fired power plants will generally require 10-15 percent additional water consumption.

An examination of the wastewater characterizations indicates that there are no significant wastewaters associated with the SO_x control systems. There are significant wastewater streams from uncontrolled power plants. The insignificant wastewater streams from SO_x control processes can be treated with developed technology to allow reuse or discharge. The effect of these streams on receiving stream water quality is expected to be negligible.

1.2 Presentation of Results

The presentation of the results of this assessment is organized in the following manner:

- Process Descriptions

Process descriptions for a power plant uncontrolled for SO_x emissions and for each process involved in the various control strategies are presented in Section 3. Process descriptions are given to allow a basic understanding of the process function, components, operating variables, and process flows. A simplified flow diagram for each process is included.

- Assessment of the Impact of SO_x Control Strategies on Coal-Fired Power Plant Water Consumption

The make-up water requirements for each of the systems and operations in the uncontrolled power plant and each of the processes involved in the various SO_x control strategies are discussed in Section 4. For purposes of discussion a base case calculation for a 500 MW power plant, burning 3.5% sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb) is used. The results of calculations for the remaining model plant systems are presented in tables at the end of Section 4. A detailed presentation of the calculation methods and assumptions is presented in Appendix B.

- Characterization of Process Wastewaters

Section 5.0 characterizes the composition and quantity of wastewaters from an uncontrolled power plant, and each of the processes involved in the various SO_x control strategies. Wastewater quality varies widely due to differences in influent water quality and the use of different water management techniques. It is therefore necessary to characterize the wastewater streams in a general manner. Characteristics of example compositions and flow rates of effluent streams are given to illustrate ranges of operation wherever possible. An assessment of relative impacts is made.

- Examination of SO_x Control System Wastewater Characteristics and Applicable Treatment Technology

A detailed examination of each of the wastewater streams for SO_x control technologies and applicable water treatment technology is presented in Section 6. This examination includes a characterization of effluent composition, an assessment of effect on receiving stream water quality, and possible water treatment technologies.

2.0 RESULTS AND CONCLUSIONS

This section presents the major findings and conclusions of the study. It summarizes the results of water consumption calculations, the characterization of process wastewaters, the assessment of effluent effects on receiving stream water quality, possible water treatment technology, and a comparison of the model plant systems. A base case 500 MW power plant burning 3.5% sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb) is used. This base case is more completely defined in Appendix A.

2.1 Results of Water Consumption Calculations

Water consumption calculations were performed to determine the additional requirement for power plants controlled by alternative SO_x control strategies. Calculations were performed for four uncontrolled power plant water systems. This was done to characterize the wide range of current practice. Calculations were also performed for each of the five FGD systems under study, the physical coal cleaning process, and the SO_2 conversion processes. In each calculation, specific requirements were determined, and their sum was taken as the process requirement. A detailed discussion of the water consumption calculations is given in Section 4.

2.1.1 Uncontrolled Power Plant Water Consumption

The results of the calculations for uncontrolled model power plant water systems are summarized in Table 2.1-1. Once-through cooling requires enormous quantities of water. The use of recirculatory cooling systems reduces this requirement significantly, even at low cycles of concentration. Ash handling systems also require large quantities of water, but

TABLE 2.1-1. BASE CASE 500 MW SYSTEM: MODEL POWER PLANT WATER CONSUMPTION

System Number	System Description	Cooling Water System		Ash Handling System		General Services Water		Boiler Makeup		Total Fresh Water Makeup Requirement	
		m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	Once-through	13	(210,000)	0.23	(3600)	0.05	(750)	0.0006	(9)	13.5	(214,000)
2	Partial Recirculatory	0.44	(7,000)	0.07	(1100)	0.05	(750)	0.0006	(9)	0.57	(9,000)
3	Recirculatory	0.32	(5,600)	0.02	(300)	0.05	(750)	0.0006	(9)	0.38	(60,000)
4	Zero Discharge	0.25	(4,000)	0	(0)	0.01	(190)	0.0006	(9)	0.25	(4,000)

[a] These power plant water systems are discussed in detail in Section 4.1

[b] The base case is a 500 MW power plant operating at an efficiency of 37%; 3.5% S coal; average heating value of 28 MJ/kg (12,000 Btu/lb).

[c] All power plant water systems are once-through; refer to Figure 4.1-1.

[d] Recirculatory cooling at 2.5 cycles of concentration, once-through ash handling, and once-through general services water; refer to Figure 4.1-2.

[e] Recirculatory cooling at 5.0 cycles of concentration, 50% recirculatory ash handling, and recycle of general service water blowdown to the ash handling system; refer to Figure 4.1-3.

[f] All power plant water systems are recirculatory; refer to Figure 4.1-4.

this requirement can be reduced through the use of cooling system blowdown as makeup, and/or recirculatory practices. The general service water requirement was estimated from plant data as being minor. A reduction credit is possible in these calculations through collection and reuse of these waters in the cooling water or ash handling systems. The boiler makeup requirement is not significant.

2.1.2 SO_x Control System Water Consumption

The results of water consumption calculations for the five FGD systems examined in this study are summarized in Table 2.1-2. The evaporative loss caused by the adiabatic saturation of the hot flue gases in the prescrubber or the absorber requires the significant fraction of the makeup to the FGD water systems in all cases. The Wellman-Lord Sulfite Scrubbing process also has a substantial requirement for makeup to the condenser cooling water system. Other large FGD water requirements are for a prescrubbing system blowdown, and occlusion of water in the solid wastes. These two requirements apply only to specific FGD systems, as indicated in the table.

Also included in Table 2.1-2 are the results of calculations of the water requirements for two SO_2 conversion processes. The Wellman-Lord Sulfite Scrubbing Process and the Magnesia Slurry Absorption Processes produce concentrated SO_2 product streams. These streams can be converted to either sulfuric acid or elemental sulfur. The calculations indicate a significant requirement for makeup to the product acid cooling water system, and an insignificant requirement for elemental sulfur production.

TABLE 2.1-2. BASE CASE 500 MW SYSTEM: FGD SYSTEM WATER CONSUMPTION[h]

Water[a] Requirement	Lime Wet Scrubbing		Limestone Wet Scrubbing		Wellman-Lord Sulfite Scrubbing		Magnesia Slurry Absorption		Double Alkali Wet Scrubbing			
									Lime Regenerant		Limestone Regenerant	
	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
Evaporative loss,	0.030	(480)	0.030	(480)	0.030	(480)	0.030	(480)	0.030	(480)	0.030	(480)
Loss in solid waste,	0.005	(80)	0.005	(80)					0.005	(72)	0.005	(79)
Prescrubber blowdown,					0.003	(54)	0.003	(54)				
Cooling water blowdown,					0.014	(220)						
Loss with solids drying,					0.002	(34)	0.003	(45)				
Loss in product SO_2 stream,					0.003	(5)						
SO_2 Conversion Requirement:												
Sulfuric acid					0.008	(130)	0.008	(130)				
Elemental Sulfur					0.002	(3)	0.0002	(3)				
Total	0.035	(560)	0.035	(560)	0.058	(923)[d]	0.024	(710)[b,d]	0.035	(552)[c]	0.035	(559)[c]
					0.050	(795)[b,e]	0.036	(580)[e]				
Liquid to gas ratio required for scrubber - m^3/Nm^3 (gal/1000 scf)	.005-.015	(35-110)	.005-.015	(35-110)	0.004	(3)[f]	.003-.004	(20-30)	.0007-.002	(5-15)	0.007-.002	(5-15)
					.002	(15)[g]	.002	(15)[g]				

[a] The base case is for a 500 MW power plant operating at an efficiency of 37%, burning 3.5% burning 3.5% sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb).

[b] This value is assumed for model plant calculations.

[c] The average of the two double alkali systems is used for model plant calculations.

[d] This is the total water requirement if sulfuric acid is produced.

[e] This is the total water requirement if sulfur is produced.

[f] Separate scrubbing loops are provided for each of 3 trays.

[g] Prescrubber.

[h] Make-up water requirement.

Not indicated in the table is the water requirement for the physical coal cleaning process. Make-up is required to replace losses due to occlusion in the solid waste, and drying losses. The makeup requirement was estimated to be 0.015 m^3/s (240 gpm) for a 500 MW power plant.

2.1.3 Conclusions

A review of Table 2.1-2 indicates that SO_x control through the use of an FGD system has water requirements ranging from 0.035 m^3/s (550 gpm) to 0.058 m^3/s (920 gpm), dependent upon the FGD system used. In comparison to the water requirement for once-through cooling shown in Table 2.1-1, this is insignificant. In comparison to the requirements of the recirculatory systems, FGD systems increase the water requirement 6 to 22 percent with the majority of the cases falling between 10 and 15 percent additional requirement.

2.2 Results of Wastewater Characterizations

Wastewaters were characterized for uncontrolled power plants, and for each of the processes involved in the alternative SO_x control strategies.

2.2.1 Uncontrolled Power Plant Wastewaters

Wastewater quality is highly dependent on the power plant water management system, the influent water quality, plant layout, and treatment practices. Because of these reasons, it was necessary to characterize the wastewaters from each source in a general manner. Typical flow rates and compositions are given in Table 2.2-1 to the extent possible. A review of Table 2.2-1 indicates that power plants are a significant source of wastewaters. The sources of largest concern are

TABLE 2.2-1. CHARACTERIZATION OF POWER PLANT WASTEWATER STREAMS

Power Plant System	Effluent Stream	Characteristics
Cooling Water	once-through discharge	The effluent water quality is essentially equivalent to that of the influent water. There may be slight differences due to the presence of corrosion products or traces of chemicals used in treatment practices.
	recirculatory system blowdown	Soluble species enter the system in the makeup water, and are concentrated to levels ranging typically from 1500 to 10,000 mg/ℓ (ppm). Intimate contact of air and water in a cooling device allows the entry of particulate matter and soluble gases into the system. Traces of chemicals used for treatments to prevent scale and corrosion may also be present. These chemicals may include inorganic polyphosphates, chelating agents, polyelectrolyte antiprecipitants, and organic/polymer dispersants for scale control, chromate, zine, phosphate, silicates, or certain proprietary organics for corrosion inhibition.
Ash Handling	bottom ash sluice water	Bottom ash forms as a fused material and therefore has little impact on sluice water quality. It has excellent settling characteristics, and is chemically inert with water.
	fly ash sluice water	Fly ash has poorer settling characteristics than bottom ash, however, low turbidities are observed with adequate residence times for sedimentation. Fly ash contains a broad spectrum of soluble inorganic salts which can be leached into the sluice water, resulting in sodium, potassium, calcium, magnesium, chloride, sulfates, etc., in solution.

(Continued)

TABLE 2.2-1 (Continued)

Power Plant System	Effluent Stream	Characteristics
Water Conditioning	sedimentation clarifier underflow	This stream has high suspended solid concentrations, and traces of coagulants and flocculants such as alum, aluminate, ferric chloride, or copperas.
	filtration backwash water	This stream has a high suspended solids content.
	Lime/lime-soda softening clarifier underflow	This stream contains traces of coagulants and flocculants such as alum, aluminate, ferric chloride, or copperas. Its hardness is typically 50 mg/ℓ as $CaCO_3$, and its pH is approximately 10.
	ion exchange regeneration waste streams	The backwashing stream from this process has a high suspended solids content. The spent regenerate has extreme pH, and contains high concentrations of ions eluted from the exchange material. The eluted ions represent the chemical species that were removed from water during the service cycle of the process.
	evaporation blowdown	Impurities in the feedwater are concentrated. Thus, the water quality of the blowdown is dependent on this influent water quality, and the degree of concentration in the evaporator. The total dissolved solids level is typically 1000 to 2000 mg/ℓ (ppm) and the pH is 9 to 11.

TABLE 2.2-1 (CONTINUED)

Power Plant System	Effluent Stream	Characteristics
Steam Generation	boiler blowdown	Boiler feedwater impurities are concentrated, and a blowdown is taken to maintain a desired level of suspended and dissolved solids. This stream contains a high level of dissolved solids, traces of corrosion products, and traces of chemicals used for treatment to prevent scale. These chemicals may include inorganic phosphates or chelating agents such as EDTA or NTA. The pH ranges from 8.0 to 9.5.
General Services	equipment cleaning and washing waste streams	These streams may have extreme pH, high suspended and dissolved solids contents, and high oxygen demands (BOD and/or COD). These streams may contain detergent constitutents.
	coal pile runoff	This stream may have a pH as low as 2 to 3, and has high suspended and dissolved solid content.

the recirculatory cooling system blowdown, ash sluicing water, and wastes from water conditioning operations. These streams have high flow rates, and contain high concentrations of various suspended and dissolved solids. Also, these streams contain trace amounts of coagulants, flocculants and metals. Some of these streams have extreme pH.

2.2.2 SO_x Control Process Wastewaters

A review of Table 2.2-2 indicates that there are no significant wastewater streams from any of the processes involved in the alternative SO_x control systems.

The Wellman-Lord and Magnesia Slurry FGD systems require a prescrubbing system blowdown. (With high chloride coals, a prescrubber may also be required for Double Alkali.) This stream has a high chlorides concentration, but otherwise has the same composition as fly ash sluice water. It has approximately 1 percent of the flow rate of the fly ash sluice water. The high chlorides concentration is usually diluted in the ash pond, before discharge to a receiving stream.

- The Wellman-Lord Sulfite Scrubbing Process and the sulfuric acid plant require a cooling water system blowdown. This blowdown is equivalent in composition to the power plant cooling system blowdown, and is 3 to 5 percent of the flow rate.

- The Double Alkali Wet Scrubbing Process and the Magnesia Slurry Absorption Process may require small purges. These purges cannot be discharged directly to a receiving stream because of their very poor quality, but water treatment technology is available. Reuse or discharge is possible after treatment.

TABLE 2.2-2. CHARACTERIZATION OF SO_x CONTROL SYSTEM PROCESS WASTEWATER STREAMS

SO_x Control Process	Effluent Stream	Characteristics
Lime/Limestone Wet Scrubbing		There are no wastewater streams associated with these processes in normal closed loop pond operation.
	Possible Scrubbing Liquor Purge	Aerospace Corporation has reported that in catastrophic situations or in operating conditions below 50 percent of design loading, a purge may be necessary. This purge would have the same composition as the scrubbing liquor (BO-203).
	Solid waste	This study does not consider the potential impact of the solid waste on surrounding water quality. Aerospace Corporation is studying this specific problem in another task in this program for the OAQPS.
Wellman-Lord Sulfite Scrubbing	Prescrubbing System Blowdown	The prescrubber blowdown may have a chloride concentration ranging from 10-20,000 mg/ℓ (ppm) and a suspended solids content of approximately 5 percent. Trace constituents of the liquor will be similar to those of fly ash sluice waters. The blowdown rate is approximately 1 percent of the fly ash sluice water requirement.
	Condenser Cooling Water System Blowdown	The quality of this stream is equivalent to the quality of the power plant cooling water system blowdown. The blowdown rate is approximately 5 percent of the power plant blowdown rate.

(Continued)

TABLE 2.2-2. (Continued)

SO_x Control Process	Effluent Stream	Characteristics
Magnesia Slurry Absorption	Prescrubbing System Blowdown	The prescrubber blowdown may have a chloride concentration ranging from 10-20,000 mg/ℓ (ppm) and a suspended solids content of approximately 5 percent. Trace constituents of the liquor will be similar to those of fly ash sluice waters. The blowdown rate is approximately 1 percent of the fly ash sluice water requirement.
	Intermittent Purge	In the developmental stages of this process an intermittent purge was taken to remove impurities that enter the system with the makeup water and makeup MgO. McGlammery et al (MC-076) have estimated that a purge rate of approximately 63 cm^3/s (1 gpm) will be necessary for a 500 MW power plant. The purge will contain 1.2 percent $MgSO_3$, 15 percent $MgSO_4$, and various trace impurities.
Double Alkali Wet Scrubbing	Possible Prescrubbing System Blowdown (Not Required in Normal Operation)	The prescrubber blowdown may have a chloride concentration ranging from 10-20,000 mg/ℓ (ppm) and a suspended solids content of approximately 5 percent. Trace constituents of the liquor will be similar to those of fly ash sluice waters. The blowdown rate is approximately 1 percent of the fly ash sluice water requirement.
	Sulfate Purge	A purge may be taken off of the clarifier supernate to remove sodium sulfate and nonsulfur/calcium solubles from the system. The purge stream has high sodium, sulfate, sulfite, and nonsulfur/calcium soluble species concentrations. The nonsulfur/calcium

(Continued)

TABLE 2.2-2. (Continued)

SO_x Control Process	Effluent Stream	Characteristics
		species are impurities which enter the system in makeup water and makeup lime or limestone.
	Solid Waste	This study does not consider the potential impact of the solid waste on surrounding water quality, as discussed under Lime/Limestone Wet Scrubbing.
Physical Coal Cleaning		Modern physical coal cleaning plant water systems operate in closed loops. There are no wastewater streams from these plants.
SO_2 Conversion Processes:		
Elemental Sulfur Production		There are no wastewater streams from this process.
Sulfuric Acid Production	Product Acid Cooling Water System Blowdown	The quality of this stream is equivalent to the quality of the power plant cooling water system blowdown. The blowdown rate is approximately 3 percent of the power plant blowdown rate.

- The Lime/limestone Wet Scrubbing Process should have no wastewater streams in normal closed loop operation. All pond water is recycled to the process. Net water gains by the pond due to rainfall can be adjusted for by cutbacks in raw water makeup to the process. However, a catastrophic condition, loading at less than 50 percent of design capacity, or operator error may require a purge. Water treatment technology is available to handle this stream to allow reuse or discharge.

2.3 Effect of SO_x Control System Wastewaters on Receiving Streams

- The scrubbing liquor purges for the Magnesia Slurry Absorption and Double Alkali Wet Scrubbing Processes should not be discharged directly to a receiving stream. Some treatment of these streams is necessary to allow reuse or discharge.

- The prescrubber blowdown is a wastewater source of high chloride concentration. The other stream components are equivalent to fly ash sluice waters. The normal treatment procedure is to route this stream to the ash pond. The chlorides are diluted to 70 mg/ℓ (ppm) in an ash pond overflow of 0.13 m^3/s (2000 gpm) for the base case. The additional impact on the receiving stream is expected to be minimal, but in site specific instances treatment of the concentrated stream may be required.

- The Wellman-Lord condenser cooling water system and the sulfuric acid plant product cooling water system require blowdowns. The blowdowns are of equivalent quality, and 3 to 5 percent of the quantity of corresponding power plant cooling water system blowdowns. Therefore, the additional impact of this wastewater on receiving stream water quality is expected to be negligible.

- Aerospace Corporation has reported that a purge of lime/limestone scrubbing slurry may be required in catastrophic situations, or in situations where these systems are operated below some critical design loading level (approximately 50 percent) (BO-203). This purge cannot be discharged directly to a receiving stream.

2.4 Examination of Water Treatment Technology Applicable to SO_x Control System Wastewater Streams

- Magnesia Slurry Absorption Process Purge - Several purge treatment techniques have been suggested by McGlammery, et al (MC-076) for the disposal and/or partial recovery of the sulfate purge. Conventional treatment techniques, such as reverse osmosis, vapor compression distillation, flash evaporation, evaporation in a deadend pond, or softening-ion exchange, could be used to treat the wastestream from the recovery processes.

- Double Alkali Sulfate Purge - Several methods for sulfate removal have been suggested by Kaplan (KA-227). If a purge is necessary to maintain a desired level of soluble nonsulfur/calcium species, the constituents remaining after sulfate removal can be removed with developed water treatment technology (i.e., reverse osmosis, vapor compression distillation, flash evaporation, or softening-ion exchange). If oxidation has been limited, the purge is discharged with the solids as occluded water.

- Prescrubber Blowdown - As recirculatory systems become predominant as a national zero discharge goal approaches, treatment of this stream will become necessary. Developed water treatment technologies that can be applied to this stream include reverse osmosis, vapor compression distillation, flash evaporation, and softening-ion exchange.

- Cooling System Blowdown - Recirculatory practices and required treatment applicable in power plant cooling systems can be applied to this stream as required.

- Possible Lime/Limestone Purge - Aerospace Corporation has reported that conventional lime-soda treatment will allow reuse of this stream within power plant water systems. If followed by a reverse osmosis treatment, the stream would be suitable for discharge and use in public water supply (BO-203). Other water treatment technologies are applicable but less favorable economically.

2.5 Comparison of Model Plant Systems

The major makeup water requirements for the five alternative SO_x control systems are given in Table 2.5-1 for a 500 MW power plant. The makeup water replaces losses due to evaporation in the absorber or prescrubber, occlusion in the solid waste, blowdown from the prescrubbing system, and blowdown of cooling water systems. The dominant makeup requirement replaces evaporative losses in the prescrubber or absorber. This loss alone demands 60 to 90 percent of the total makeup for the various FGD systems. Large amounts of makeup water are required for the physical coal cleaning process, and for makeup to cooling water systems in specific SO_x control systems.

TABLE 2.5-1. SO_x CONTROL SYSTEM MAKEUP WATER REQUIREMENTS[a]

	m^3/s	gpm
Evaporative loss	0.030	480
Cooling water system:		
Wellman-Lord Sulfite Scrubbing	0.014	220
Sulfuric Acid Production	0.008	130
Physical Coal Cleaning	0.015	240
Prescrubber Blowdown	0.003	54
Occlusion in solid wastes	0.005	75

[a] The example requirements given in table are for a base case 500 MW power plant.

An examination of the model plant system calculations (presented in Section 4.9) allows the following conclusions:

- In general, for a given size power plant and percent sulfur removal, the water requirements for the five FGD systems increase in the following order:

Double Alkali < Lime < Limestone < Magnesia Slurry < Wellman-Lord.

If a prescrubber is used with Double Alkali, its water requirements will be greater than those of the Limestone process.

- The water requirement increases in direct proportionality with the power plant size.

- In nonregenerable FGD systems, increased sulfur removal increases the amount of water occluded in the solid waste by direct proportion. Doubling the amount of sulfur removed, however, increases the total makeup requirement on the order of 10 to 15 percent.

- In control systems using physical coal cleaning, and reduced SO_2 removal in the FGD system, water consumption is increased by the requirement for the coal cleaning process. The major FGD loss is due to evaporation in the scrubber, and this loss is constant regardless of the initial coal sulfur content.

- Increasing the amount of sulfur removal in the FGD system does not significantly alter the makeup requirement due to the dominant evaporative loss.

- In control systems using low sulfur coals, a large increase in water consumption is seen if an FGD system is required. FGD system removal of 50 and 90 percent of the sulfur content require essentially the same amounts of water due to the dominant evaporative losses. Burning low sulfur coal without an FGD requires no makeup water. FGD applications to 500 MW power plants burning low sulfur coal require 0.035 m^3/s (550 gpm) makeup.

3.0 PROCESS DESCRIPTIONS

This section presents process descriptions for the water system of a coal-fired power plant uncontrolled for SO_x emissions, and for each process involved in the alternative SO_x control systems. These descriptions provide a basic understanding of the process, system components, operating variables, and process flows. A simplified flow diagram for each process is included. Descriptions are included for the following processes:

- Coal-fired Power Plant Water System: Uncontrolled for SO_x Emissions,
- Lime Wet Scrubbing,
- Limestone Wet Scrubbing,
- Wellman-Lord Sulfite Scrubbing,
- Magnesia Slurry Absorption,
- Double Alkali Wet;
- Physical Coal Cleaning, and
- SO_2 Conversion.

3.1 Coal-Fired Power Plant Water System: Uncontrolled for SO_x Emissions

A fossil-fuel fired power generating station uses large quantities of water. The primary use is for the condensation of exhaust steam from turbines in the condenser cooling system. Other power plant water uses include steam generation, ash handling, water conditioning, cleaning operations, and miscellaneous operations.

The major consumers of water in a power generating station unequipped with an FGD system are the condenser cooling system and the ash handling system. There are several process variations of each of these two systems. However, all can be classified as either once-through systems or recirculatory systems. The quantities of water required/discharged in the recirculatory systems are much less than those for once-through systems. The circulating water flow rate in the recirculatory system is comparable to the discharge flow rate in the once-through system. Although the water requirement is less in recirculatory systems, impurities are concentrated, and may cause scale formation. A blowdown stream is taken from the system to avoid excessive buildup of dissolved species.

Other power plant processes which use and consume water are steam generation, water conditioning, cleaning operations, and miscellaneous operations. The water consumption in these processes, however, is insignificant in comparison to the consumption in the condenser cooling system.

3.1.1 Process Description

Water management in a fossil-fuel power generating station must consider the following water consumptive processes and operations:

1) Condenser Cooling System
2) Ash Handling System
3) Steam Generating System
4) Water Conditioning Operations
5) Cleaning Operations
6) Miscellaneous Operations.

Figures 3.1-1 and 3.1-2 show simplified flow diagrams for power plant water systems which use recirculatiory and once-through processes, respectively, in both the condenser cooling and ash handling systems.

Condenser Cooling System

Approximately 45 percent of a fossil-fuel-fired generating station's energy is removed and ultimately discharged to the environment by the condenser cooling system. Basically, two condenser cooling systems are employed by the electric utility industry: 1) once-through system and 2) recirculating system.

In once-through cooling systems, the total cooling water flow for heat removal is discharged as wastewater effluent. After passing through the condenser, the cooling water is discharged to a heat sink, *e.g.*, a river, lake, or pond, where the heat is dissipated. Thermal pollution of the heat sink is the major problem associated with once-through cooling systems.

Recirculating cooling systems employ cooling devices such as cooling towers, spray ponds, canals, etc., which allows the reuse of recirculated cooling water. These devices promote cooling primarily by evaporating a portion of the recirculating water flow. Thus, impurities that come into the system through makeup water or other sources are concentrated. A blowdown stream is withdrawn from the system to control the concentration of impurities. The quantity of blowdown is set by the maximum concentration of a limiting impurity, *e.g.*, dissolved solids, that can be tolerated in the system, or by the solubility limit of scaling salts such as calcium carbonate or sulfate.

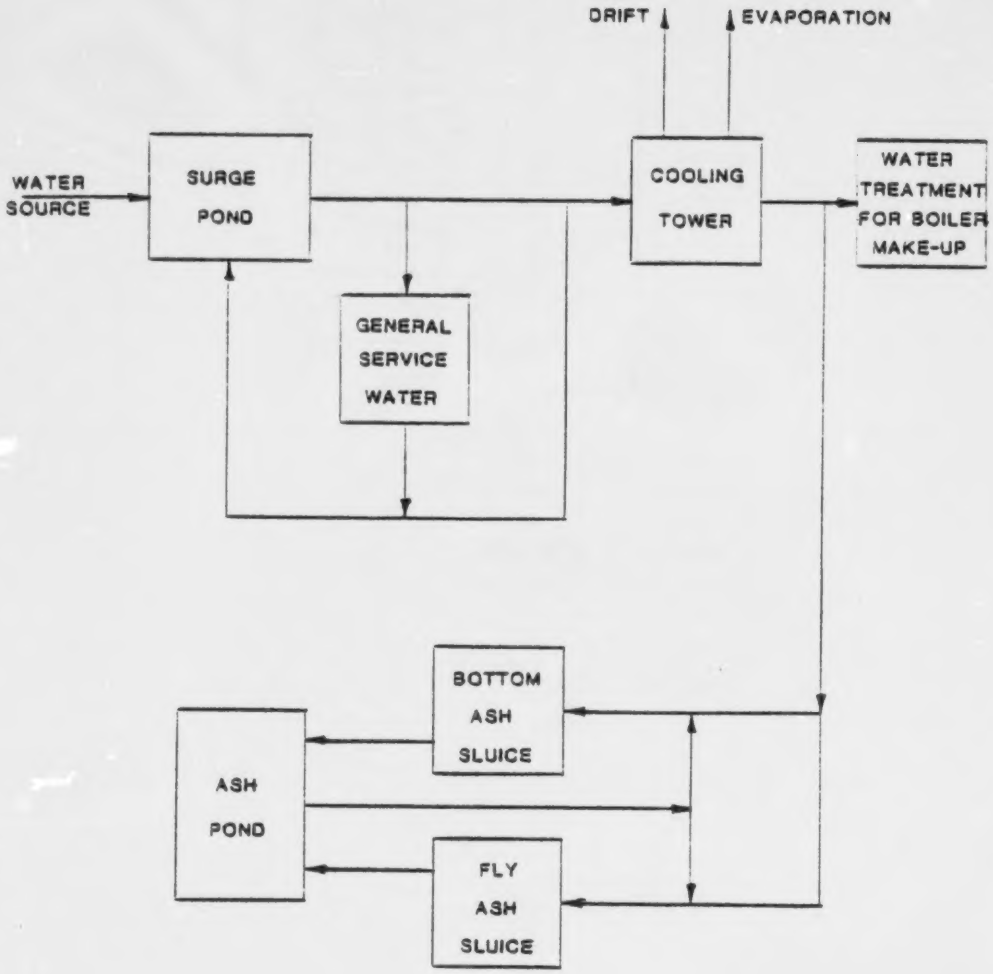
DRIFT
EVAPORATION
WATER SOURCE
SURGE POND
COOLING TOWER
WATER TREATMENT FOR BOILER MAKE-UP
GENERAL SERVICE WATER
BOTTOM ASH SLUICE
ASH POND
FLY ASH SLUICE

Figure 3.1-1. Power plant water system: recirculatory water management.

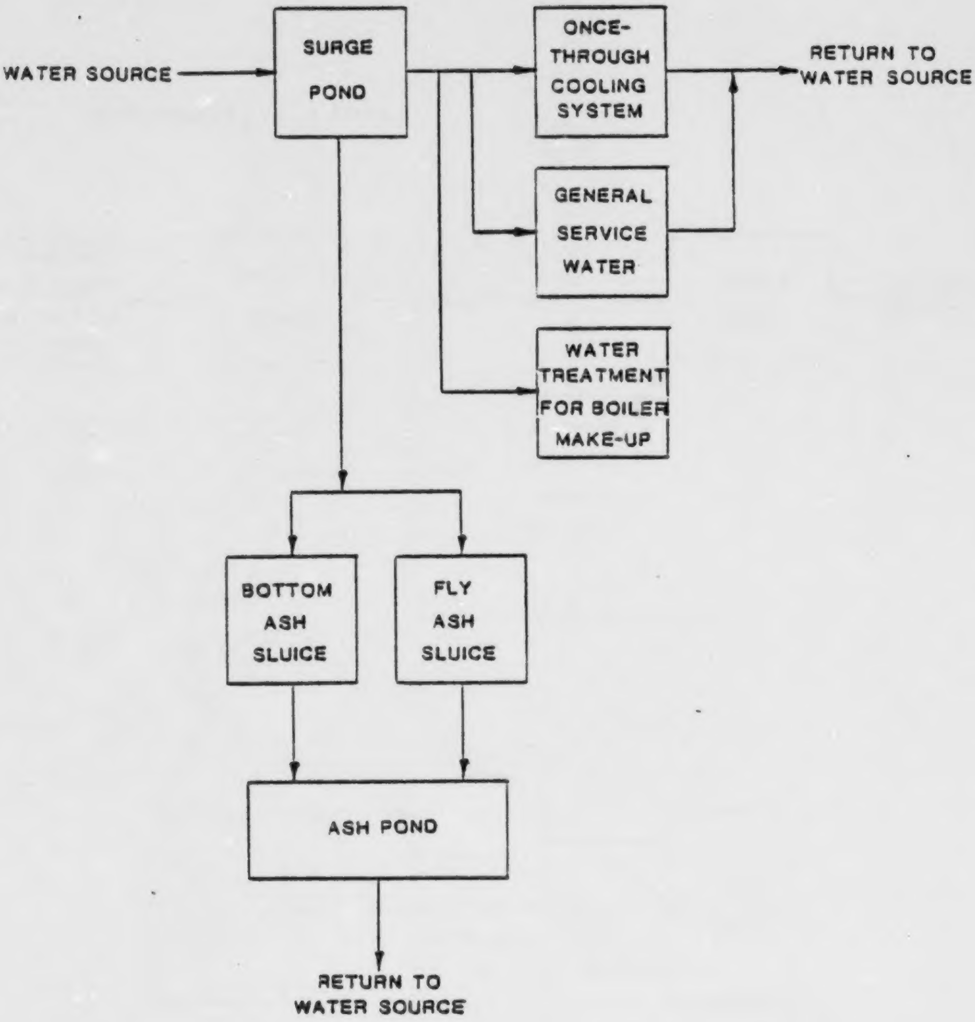
WATER SOURCE
SURGE POND
ONCE-THROUGH COOLING SYSTEM
RETURN TO WATER SOURCE
GENERAL SERVICE WATER
WATER TREATMENT FOR BOILER MAKE-UP
BOTTOM ASH SLUICE
FLY ASH SLUICE
ASH POND
RETURN TO WATER SOURCE

Figure 3.1-2. Power plant water system: once-through water management.

Ash Handling

Ash, a solid by-product of coal combustion, is formed in a power plant boiler as bottom or fly ash. Bottom ash settles in the boiler firebox and must be removed from the boiler in order to maintain system operability. Fly ash is entrained with the flue gas, leaves the boiler as particulate matter, and is normally collected in flue gas cleaning equipment. The conveyance of both bottom ash and fly ash to points of disposal constitutes ash handling. Ash handling systems employ either pneumatic or hydraulic mechanisms for ash transportation. Hydraulic (wet sluicing) systems produce wastewater streams.

Coal-fired generating stations require formal ash handling facilities due to the quantity of ash produced during coal combustion. The ash content of U.S. coals range from 6 to 20 weight percent. The average value is approximately 11 weight percent (EN-127). The distribution between bottom ash and fly ash is greatly influenced by boiler furnace design and operating mode. Ash distribution can affect the water balance for a hydraulic ash handling system. The chemical differences between fly and bottom ashes can also affect sluicing water quality.

Bottom ash generally forms as a fused, clinker-type material and is removed by wet sluicing to the ash pond. Hydraulic design considerations dictate a sluice water of 5 to 10 percent solids. In practice, slurries of less than 1 percent solids are used depending on such factors as plant design, location, and operating circumstances (AY-007).

Fly ash can be collected in the dry form by cyclones, fabric filters, dry electrostatic precipitators, etc., and in a water slurry by wet scrubbers, wet electrostatic precipitators, etc. Fly ash, collected in either the wet or dry form, is commonly sluiced to ash ponds for sedimentation of the suspended fly ash solids. The minimum sluice water requirement is set by hydraulic design considerations. For fly ash, sluice water compositions range from 5 to 10 percent solids. However, as with bottom ash sluicing, sluice water compositions of less than 1 percent solids are used in certain cases (AY-007).

In a once-through ash sluicing system, both bottom and fly ashes are sluiced to a disposal pond where the ash settles. The overflow from the disposal pond is then discharged. In a recirculating system, a portion or all of the pond overflow is recycled to the system. Scale problems may result if soluble salts are leached from the fly ash and concentrated in the recirculating system. This may be avoided by treating a slipstream with lime or soda-ash softening.

Steam Generation

Power plant boilers are either of the once-through or the drum-type design. Once-through designs are employed exclusively in high pressure, super-critical boilers. No wastewater streams are produced by operation of once-through systems associated with their operation. Drum-type boilers operate at subcritical conditions. Steam generated in the drum-type units is in equilibrium with liquid boiler water. Boiler water impurities are, therefore, concentrated in the liquid phase as steam is generated in these units. These impurities are ultimately removed in a liquid blowdown stream.

Water Conditioning Operations

Water requirements for conditioning operations include water required for backwashing filters in lime softening processes and regenerating demineralizer resins. Some degree of water treatment is practiced in all power plants. Water is treated primarily to remove suspended solids and/or dissolved salts. Sedimentation and filtration are used for removing suspended solids. Lime/lime-soda softening, ion exchange, and evaporation are used to remove dissolved solids. Water conditioning schemes employ these basic processes singly or in multiple combinations.

Cleaning Operations

Heat transfer surfaces in the boiler and steam condenser are chemically cleaned prior to plant start-up. In addition, operational cleaning occurs during the plant's service life. Operational cleaning removes scale and corrosion products that accumulate on the boiler's steam-side and on the water-side of the steam condenser. The frequency of chemical cleaning varies from power plant to power plant. The active reagents in cleaning solutions are acidic or alkaline in nature depending on the deposits to be removed. Ninety percent of cleaning operations employ acidic solutions. Acid solutions attack all forms of alkaline scale (*i.e.*, $CaCO_3$, $Mg(OH)_2$, etc.), silica scale, and corrosion deposits containing iron. The majority of these compounds contain hydrochloric acid in solution strengths ranging from 5.0 to 7.5 percent (AY-007). Alkaline solutions are employed to remove deposits passive to acid attack and to neutralize residuals resulting from acid cleanings.

Miscellaneous Operations

A number of miscellaneous operations also constitute sources of plant wastewater. For the most part, the impact of wastewater from these miscellaneous operations is small compared to those discussed previously; their nature is highly varied depending upon specific plant characteristics. Three typical miscellaneous operations are auxiliary cooling systems, water intake screen washings, and laboratory and sampling operations.

3.2 Lime Wet Scrubbing Process

In the lime flue gas desulfurization process SO_2 is removed from the flue gas by wet scrubbing with a slurry of calcium oxide (lime). The principal reaction for absorption of SO_2 by the lime slurry is:

$$SO_{2(g)} + CaO_{(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)}. \qquad (3.2\text{-}1)$$

Oxygen absorbed from the flue gas or surrounding atmosphere causes the oxidation of absorbed SO_2. The calcium sulfite formed in the principal reaction and the calcium sulfate formed through oxidation are precipitated as crystals in a hold tank. The crystals are recovered in a solid/liquid separator. Waste solids disposal is accomplished by ponding or landfill. The clear liquor is recycled.

3.2.1 Process Description

The design of a lime wet scrubbing system includes the following process areas:

1) SO_2 Absorption,

2) Solids Separation, and

3) Solids Disposal.

Figure 3.2-1 shows a generalized flow diagram for the lime wet scrubbing process.

SO_2 Absorption

SO_2 is removed from flue gas in a wet scrubber by absorption into a circulating slurry of lime. Calcium sulfite is

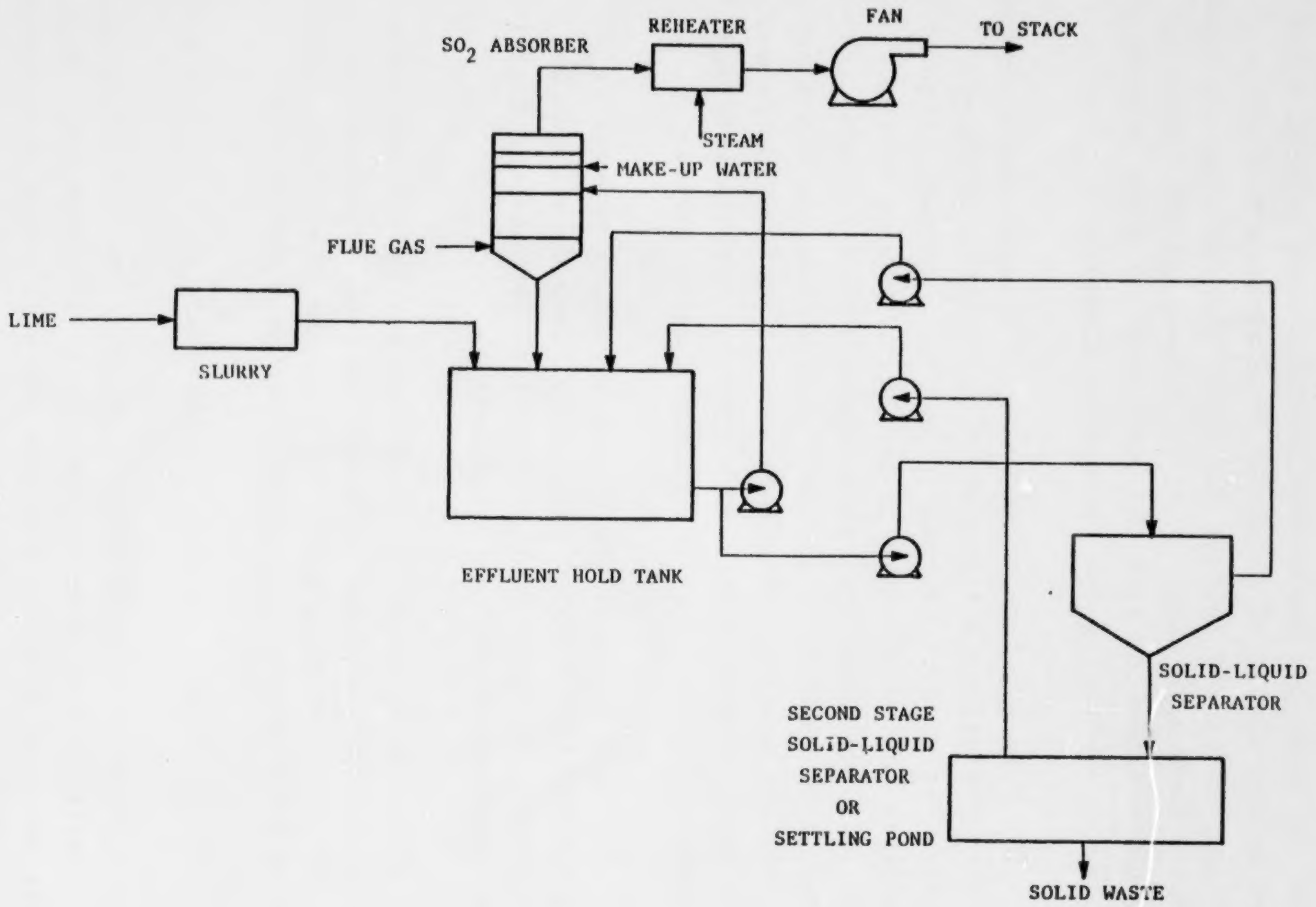
SO_2 ABSORBER
REHEATER
FAN
TO STACK
STEAM
MAKE-UP WATER
FLUE GAS
LIME
SLURRY
EFFLUENT HOLD TANK
SOLID-LIQUID SEPARATOR
SECOND STAGE SOLID-LIQUID SEPARATOR OR SETTLING POND
SOLID WASTE

Figure 3.2-1. Process flow diagram for the lime wet scrubbing process.

formed in the principal absorption reaction, and calcium sulfate is formed as the result of oxidation in a secondary reaction. The calcium sulfite and sulfate formed in the scrubber are then precipitated in a hold tank. A 10-15 percent solids slurry is recycled to the absorber from the hold tank. A bleed stream is sent to solids dewatering for subsequent disposal.

The flue gas can be pretreated for particulate removal with an electrostatic precipitator or particulate scrubber. Particulates can also be removed in the SO_2 absorber, although this increases the solids load in the SO_2 scrubbing system. In addition, it is believed that some components of fly ash catalyze the oxidation of sulfite to sulfate, thus increasing the potential for sulfate scaling. The selection of particulate removal method is based on assessments of operational reliability and the economics of installing particulate control devices.

The feed material for a lime scrubbing process is usually produced by calcining limestone. Carbide sludge, an impure, slaked lime by-product of acetylene production, has also been used successfully at two installations.

The absorption of SO_2 from flue gas by a lime slurry constitutes a multiphase system involving gas, liquid and solid phases. The reaction of gaseous SO_2 with the lime slurry yielding calcium sulfite hemi-hydrate is shown in Equation 3.2-1.

$$SO_{2\,(g)} + CaO_{(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)}. \qquad (3.2\text{-}1)$$

The solid sulfite is only slightly soluble in the scrubbing liquor and will precipitate to form an inert solid for disposal. Some CO_2 may be absorbed from the flue gas and will react in a similar manner to form solid calcium carbonate.

In most cases some oxygen will also be absorbed from the flue gas or surrounding atmosphere. This leads to oxidation of absorbed SO_2 and precipitation of solid $CaSO_4 \cdot 2H_2O$. The reaction for this step is:

$$SO_{2\,(g)} + \tfrac{1}{2}O_{2\,(g)} + CaO_{(s)} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O_{(s)}. \qquad (3.2\text{-}2)$$

The extent of oxidation can vary considerably, from almost zero to 40 percent. In some systems treating dilute SO_2 flue gas streams, sulfite oxidation as high as 90 percent has been observed. The mechanism for sulfite oxidation is not completely understood. The rate is known to be a strong function of oxygen concentration in the flue gas and liquor pH. It may also be increased by trace quantities of catalysts in fly ash entering the system.

Several types of gas-liquid contactors can be used as the SO_2 absorber. These differ in SO_2 removal efficiency as well as operating reliability. Four types of contactors are generally used for SO_2 removal:

- venturi scrubbers,
- spray towers (horizontal and vertical),
- grid towers, and
- mobile bed absorbers (such as TCA marble bed and turbulent contact absorber).

The liquid to gas ratio (L/G) typically ranges between 0.005-0.015 m^3/Nm^3 (35-110 gal/1000 scf) depending upon the type of contactor. Simple impingement devices are placed downstream from the absorber to remove mist entrained in the flue gas.

The absorber effluent is sent to a hold tank for precipitation of calcium sulfite and sulfate. The tank is equipped with an agitator to prevent settling of solids and to maintain uniform effluent composition. Additional streams entering the tank include settling pond water, clarifier overflow, and makeup lime slurry. The hold tank is sized to allow sufficient residence time for dissipation of supersaturation and precipitation of calcium sulfite and sulfate. Too little residence time in the hold tank can cause scaling as a result of nucleation of calcium sulfite and sulfate solids in the scrubber.

Solids Separation

A bleed stream is taken off the effluent hold tank to be dewatered. This step, necessary to minimize the land area needed for sludge disposal, varies depending on the application and type of disposal.

For systems with on-site pond disposal, solids may be pumped directly from the effluent hold tank to the pond area. Clean overflow liquor from the pond would then be returned to the system. If necessary, a thickening device such as a clarifier or centrifuge can be used to increase the solids content to a maximum of about 40 weight percent. Additional dewatering to 60-70 percent solids can sometimes be achieved by vacuum filtration.

Solids Disposal

The lime flue gas desulfurization process is a non-regenerative or "throwaway" process. Sludge disposal is one of the main disadvantages of "throwaway" FGD systems as compared to "recovery" processes. The quantity of sludge produced is large in weight and volume; it requires a large waste pond or landfill area for disposal.

On-site disposal is usually accomplished by sending the waste solids to a large pond, where settling of the solids occurs. Pond water is recycled to the process hold tank for reuse.

Stabilization methods are currently under development to convert the sludge to structurally-stable, leach-resistant landfill material. When on-site disposal is not possible, the stabilized material could be trucked to an off-site landfill.

3.3 Limestone Wet Scrubbing Process

In the limestone flue gas desulfurization process, SO_2 is removed from the flue gas by wet scrubbing with a slurry of calcium carbonate. The principal reaction for absorption of SO_2 is:

$$SO_{2\,(g)} + CaCO_{3\,(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)} + CO_{2\,(g)}. \quad (3.3\text{-}1)$$

Oxygen absorbed from the flue gas or surrounding atmosphere may cause the oxidation of absorbed SO_2. The calcium sulfite formed in the principal reaction and the calcium sulfate formed via oxidation are precipitated as crystals in a hold tank. The crystals are then sent to a solid/liquid separator where the solids are removed. Waste solids disposal is accomplished by ponding or landfill. The clear liquor is recycled.

3.3.1 Process Description

The design of a limestone wet scrubbing system can be divided into the following process areas:

1) SO_2 Absorption,
2) Solids Separation, and
3) Solids Disposal.

Figure 3.3-1 shows a generalized flow diagram for the limestone wet scrubbing process.

SO_2 Absorption

SO_2 is removed from flue gas in a wet scrubber by absorption into a circulating slurry of calcium carbonate. Calcium sulfite is formed in the principal absorption reaction; calcium sulfate is formed as the result of oxidation in a secondard reaction. The calcium sulfite and sulfate formed in the scrubber are then precipitated in a hold tank. A 10-15 percent solids slurry is recycled to the absorber from the hold tank. A bleed stream is sent to solids dewatering for subsequent disposal.

The flue gas can be pretreated for particulate removal with an electrostatic precipitator or particulate scrubber. Particulates can also be removed in the SO_2 absorber, although this increases the solids loading in the SO_2 scrubbing system. In addition, it is believed that some components of fly ash catalyze the oxidation of sulfite to sulfate, thus increasing the potential for sulfate scaling. The selection of particulate removal method is based on assessments of operational reliability and the economics of installing particulate control devices.

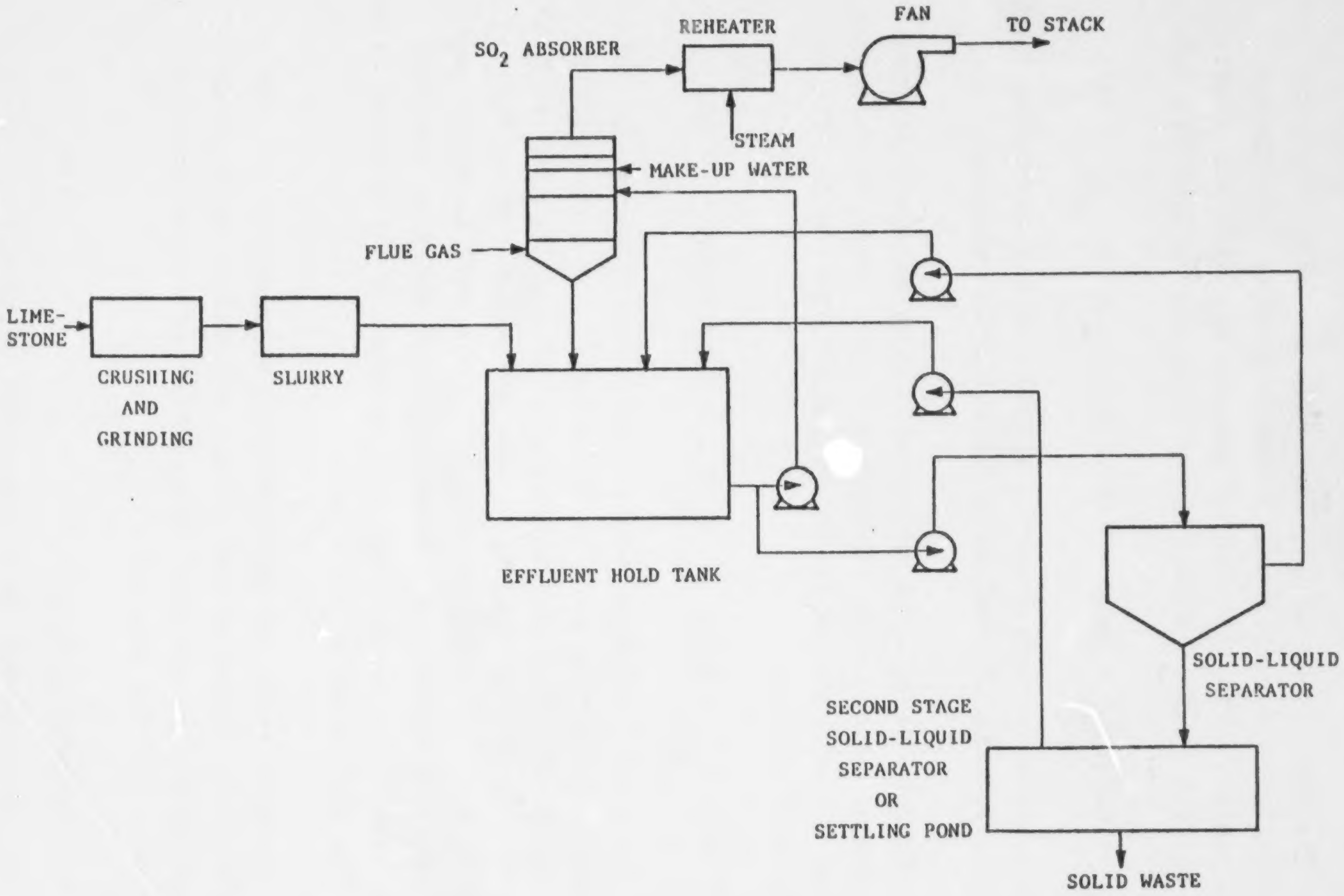
SO$_2$ ABSORBER
REHEATER
FAN
TO STACK
STEAM
MAKE-UP WATER
FLUE GAS
LIME-
STONE
CRUSHING
AND
GRINDING
SLURRY
EFFLUENT HOLD TANK
SOLID-LIQUID
SEPARATOR
SECOND STAGE
SOLID-LIQUID
SEPARATOR
OR
SETTLING POND
SOLID WASTE

Figure 3.3-1. Process flow diagram limestone wet scrubbing process.

The absorption of SO_2 from flue gas by a limestone slurry constitutes a multiphase system involving gas, liquid, and solid phases. The reaction of gaseous SO_2 with the slurry yielding calcium sulfite hemi-hydrate is shown in Equation 3.3-1:

$$SO_{2\,(g)} + CaCO_{3\,(s)} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O_{(s)} + CO_{2\,(g)}. \qquad (3.3\text{-}1)$$

The solid sulfite is only slightly soluble in the scrubbing liquor and will precipitate to form an inert solid for disposal.

In most cases some oxygen will also be absorbed from the flue gas or surrounding atmosphere. This leads to oxidation of absorbed SO_2 and precipitation of solid $CaSO_4 \cdot 2H_2O$. The reaction for this step is:

$$SO_{2\,(g)} + \tfrac{1}{2}O_{2\,(g)} + CaCO_{3\,(s)} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O_{(s)} + CO_{2\,(g)}. \qquad (3.3\text{-}2)$$

Although the extent of oxidation can vary considerably, it normally ranges from almost zero to 40 percent. In some systems treating dilute SO flue gas streams, sulfite oxidations as high as 90 percent have been observed. The mechanism for sulfite oxidation is not completely understood. The rate is known to be a strong function of oxygen concentration in the flue gas and liquor pH. It may also be increased by trace quantities of catalysts in fly ash entering the system.

Several types of gas-liquid contactors can be used as the SO_2 absorber. These differ in SO_2 removal efficiency as well as operating reliability. Four types of contactors are generally used for SO_2 removal:

- venturi scrubbers,
- spray towers (horizontal and vertical),

- grid towers, and
- mobile bed absorbers (such as TCA marble bed and turbulent contact absorber).

The liquid to gas ratio (L/G) typically ranges between 0.005 - 0.015 m^3/Nm^3 (35-110 gal/1000 scf) depending upon the type of contactor. Simple impingement devices are placed downstream from the absorber to remove mist entrained in the flue gas.

The absorber effluent is sent to the hold tank for precipitation of calcium sulfite and sulfate. The tank is equipped with an agitator to prevent settling of solids and to maintain uniform effluent composition. Additional streams entering the tank include settling pond water, clarifier overflow, and makeup lime slurry. The hold tank is sized to allow sufficient residence time for dissipation of supersaturation and precipitation of calcium sulfite and sulfate. Too little residence time in the hold tank can cause scaling as a result of nucleation of calcium sulfite and sulfate solids in the scrubber, resulting in scaling.

Solids Separation

A bleed stream is taken off the effluent hold tank to be dewatered. This step, necessary to minimize the land area needed for sludge disposal, varies depending on the application and type of disposal.

For systems with on-site pond disposal, solids may be pumped directly from the effluent hold tank to the pond area. Clean overflow liquor from the pond would then be returned to the system. If necessary, a thickening device such as a clarifier or centrifuge can be used to increase the solids content to a maximum of about 40 weight percent. Additional dewatering to 60-70 percent solids can sometimes be achieved by vacuum filtration.

Solids Disposal

The limestone flue gas desulfurization process is a non-regenerative or "throwaway" process. Sludge disposal is one of the main disadvantages of "throwaway" FGD systems as compared to "recovery" processes. The quantity of sludge produced is large in weight and volume, and requires a large waste pond or landfill area for disposal.

On-site disposal is usually accomplished by sending the waste solids to a large pond where settling of the solids occurs. Pond water is recycled to the process hold tank for reuse.

Stabilization methods are currently under development to convert the sludge to structurally-stable, leach-resistant, landfill material. When on-site disposal is not possible, the stabilized material could be trucked to an off-site landfill.

3.4 Wellman-Lord Sulfite Scrubbing Process

The Wellman-Lord Sulfite Scrubbing Process is a regenerable flue gas desulfurization process marketed by Davy Powergas. It is based on the ability of a sodium sulfite solution to absorb SO_2 and form a solution of sodium bisulfite. The sodium bisulfite solution can be thermally regenerated to produce a concentrated stream of SO_2 and the original sodium sulfite solution. The concentrated SO_2 stream can be processed to produce elemental sulfur, sulfuric acid, or liquid SO_2. The regenerated sodium sulfite solution is recycled to the absorber. In the absorption phase of the process, sulfates formed by oxidation of sulfites are removed from the system in a purge of sodium sulfate and sulfite solids.

3.4.1 Process Description

The Wellman-Lord Process consists of five processing areas:

1) Gas pretreatment,
2) Absorption,
3) Regeneration,
4) Purge treatment, and
5) SO_2 conversion.

A simplified process flow sheet appears in Figure 3.4-1. The gas pretreatment and absorption sections are essentially the same as those found in most aqueous scrubbing systems. No unique equipment is used in any of the processing areas with the possible exception of the SO_2 conversion step, which is licensed technology.

Gas Pretreatment

Flue gas is pretreated in a venturi or tray-type prescrubber to cool and humidify the gas, and to reduce fly ash and chlorides. The humidification and cooling step prevents the evaporation of excessive amounts of water in the absorber. The potential for scaling and plugging problems is reduced by the removal of fly ash; in a well designed prescrubber, 99 percent of the chlorides can be removed, thus reducing the potential for stress corrosion.

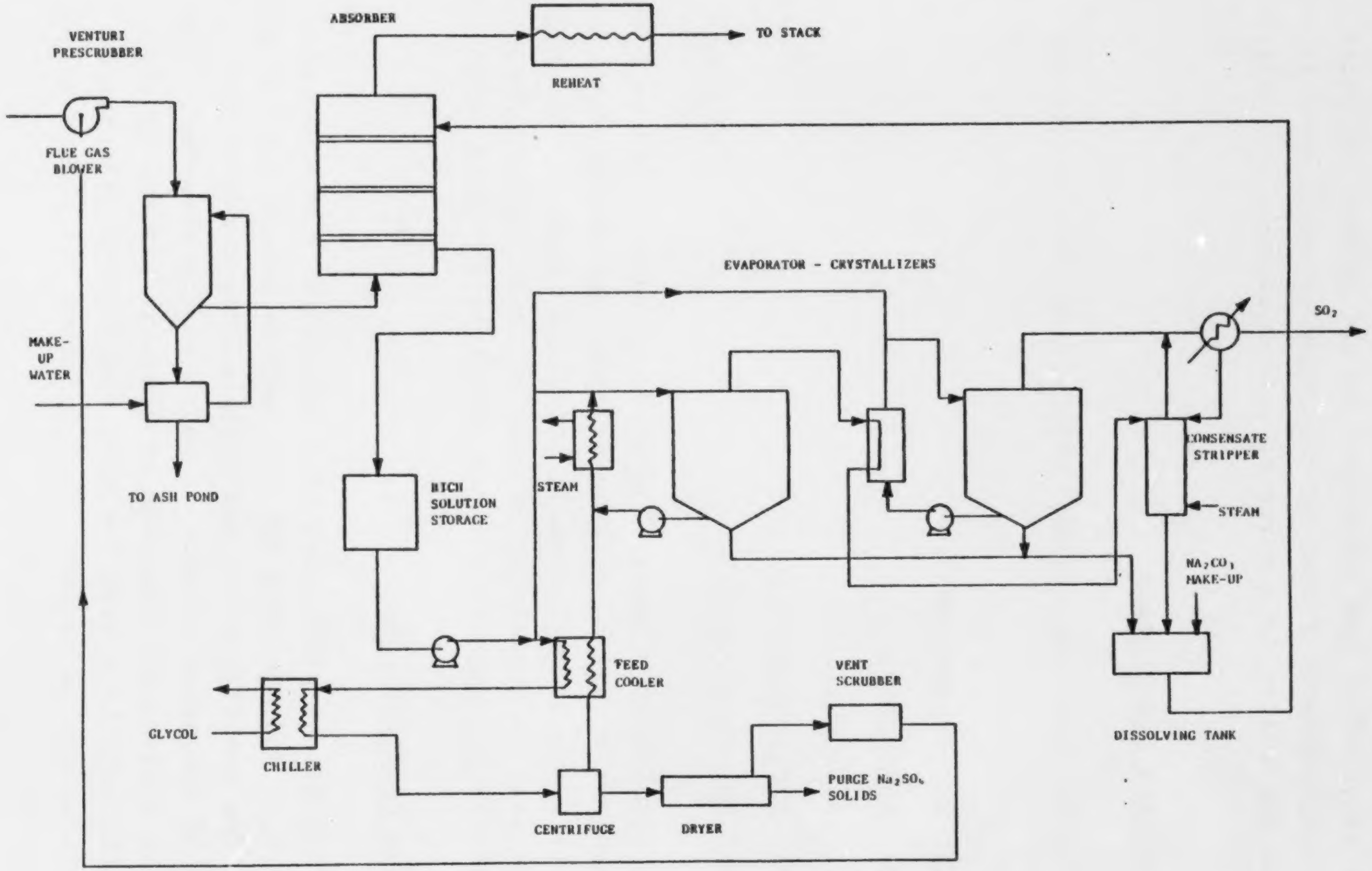
VENTURI PRESCRUBBER
FLUE GAS BLOWER
MAKE-UP WATER
TO ASH POND
ABSORBER
REHEAT
TO STACK
RICH SOLUTION STORAGE
EVAPORATOR - CRYSTALLIZERS
STEAM
FEED COOLER
GLYCOL
CHILLER
CENTRIFUGE
DRYER
VENT SCRUBBER
PURGE Na_2SO_4 SOLIDS
SO_2
CONSENSATE STRIPPER
STEAM
NA_2CO_3 MAKE-UP
DISSOLVING TANK

Figure 3.4-1. Process flow diagram for the Wellman-Lord sulfite scrubbing process.

Flue gas, exiting the electrostatic precipitator at a temperature of 150°C (300°F), is passed through a venturi or tray-type prescrubber. There the gas is cooled to around 55°C (130°F) and humidified. The venturi is preferred because 70 to 80 percent of the remaining fly ash and 95 to 99 percent of the chlorides are removed. Although a tray-type prescrubber is satisfactory for cooling and humidifying the gas with low pressure drop, it provides lower reductions in fly ash and chlorides. The fly ash and other solids collected by the prescrubber are pumped to the ash disposal pond as a slurry of approximately 5 percent solids. Since absorption of chlorides and some SO_2 and SO_2 can cause the slurry to become fairly acidic, neutralization is accomplished with lime when necessary.

Absorption

Cooled and humidified gas from the prescrubber passes upward through an absorption tower, where SO_2 is removed by absorption into the sodium sulfite scrubbing solution. After the cleaned gas is reheated to about 80°C (175°F) (so that it has the proper dew point and buoyancy), it is then exhausted to the atmosphere. The scrubbing solution is sent to regeneration and purge treatment.

Davy Powergas offers two types of absorption units: a packed tower for small volume applications and a valve tray tower for large volume applications. The valve tray unit, built in a square configuration, includes three to five trays depending on the inlet SO_2 concentration and the degree of desulfurization required. Three to four absorbers would be used in a 500 Mw installation. Because sodium sulfite has a large capacity to absorb SO_2, the feed liquor flow rate is low. Recirculation is practiced on each stage to maintain good hydraulic characteristics.

With recirculation, the L/G ratio is kept at approximately .0004 m^3/Nm^3 (3 gal/1000 scf) per tray.

The absorption of SO_2 proceeds according to Equation 3.4-1:

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (3.4\text{-}1)$$

Makeup sodium carbonate also reacts with SO_2 in the absorber to form sodium sulfite by Equation 3.4-2:

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2 \qquad (3.4\text{-}2)$$

A very important side reaction is the oxidation of sulfite to sulfate by oxygen in the flue gas as in Equation 3.4-3:

$$Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4 \qquad (3.4\text{-}3)$$

Some sodium sulfate is also formed by absorption of SO_2 from the flue gas as in Equation 3.4-4.

$$2Na_2SO_3 + SO_3 + H_2O \rightarrow 2NaHSO_3 + Na_2SO_4 \qquad (3.4\text{-}4)$$

Cooled and humidified gas from the prescrubber is passed up through the absorption tower, where the SO_2 level is reduced by at least 90 percent. The cleaned gas is reheated by heat exchange with high-pressure steam and exhausted to the atmosphere. Although alternatives to this exist, the use of steam allows coal to be used indirectly rather than premium fuels such as oil or natural gas.

Regeneration

Regeneration of sodium sulfite is accomplished by reversing the absorption reaction through the addition of heat. The absorber product and purge centrate (discussed in the following section) enter a double effect evaporator where SO_2 and water vapor are driven off. The SO_2 and water vapor are subjected to partial condensation to remove most of the water and produce a concentrated SO_2 stream. The evaporator bottoms high-solids sodium sulfite solution, condensate, and makeup Na_2CO_3 are mixed in a dissolving tank and recycled to the absorber.

The regeneration of sodium sulfite proceeds according to Equation 3.4-5:

$$2NaHSO_3 \xrightarrow[\Delta]{} Na_2SO_3 + H_2O + SO_2 \qquad (3.4\text{-}5)$$

Because of the higher temperatures, there is an increased formation of thiosulfate by the disproportionation reactions 3.4-6 and 3.4-7:

$$6NaHSO_3 \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + 2SO_2 + 3H_2O \qquad (3.4\text{-}6)$$

$$2NaHSO_3 + 2Na_2SO_3 \rightarrow 2Na_2SO_4 + Na_2S_2O_3 + H_2O \qquad (3.4\text{-}7)$$

The combined stream of absorber product liquor and purge centrate is split between the two evaporator effects, each of which operates under a vacuum. Fifty-five percent goes to the first effect, with 45 percent going to the second. The first effect operates at 95°C (200°F) and is heated with low pressure steam by an external shell and tube exchanger. The SO_2 and water vapor driven off overhead from the first effect are used to heat the second effect which operates at about 75°C (170°F). The approximately

45 percent undissolved solids content in each effect is primarily sodium sulfite. The regeneration reaction is limited by the equilibrium concentration of sulfite ion in solution. Fortunately, since sodium sulfite is less soluble than sodium bisulfite, it is continuously removed from solution by crystallization, thus driving the reaction forward. The evaporator product is sent to the dissolving tank.

The SO_2 and water vapor overhead from the evaporators is subjected to partial condensation to remove most of the water and concentrate the SO_2. The condensate, containing several hundred ppm of dissolved SO_2, is steam-stripped to lower this value. Along with a small amount of makeup water and sodium carbonate, the stripped condensate is sent to the dissolving tank. There it is agitated with the sodium sulfite slurry from the evaporators to provide absorber feed. The SO_2 stream exiting the condenser contains only 5-10 percent water. It is compressed and sent to an SO_2 conversion process.

Purge Treatment

A sidestream of the absorber product liquor is drawn off to the purge treatment area for the removal of sodium sulfate in a chiller/crystallizer.

About 15 percent of the absorber product liquor is sent to the purge treatment area for removal of sodium sulfate. This stream is precooled by heat exchange with the cold, sulfate-free product. When the purge is cooled to 0°C (32°F) in a chiller/chrystallizer, a mixture of sodium sulfate and sulfite is crystallized out. The sodium sulfate and sulfite slurry is centrifuged to produce a 40 percent solids cake. Just prior to steam drying, a secondary purge stream is drawn off the evaporators and added to the cake. This secondary purge removes thiosulfate formed in the regeneration phase of the process. The resulting product is a crystalline mixture of anhydrous sodium sulfate (70 percent), and sodium sulfite (30 percent), with small amounts of thiosulfates, pyrosulfites, and chlorides. The sulfate-free supernatant liquor from the centrifuge is heated by passing it through the feed cooler. It is then returned to the product liquor stream entering the evaporator loop. The vent gases from the dryer are cleaned to remove dust and returned to the main flue gas stream before it enters the prescrubber.

SO_2 Conversion

Several processes can be used to convert the concentrated SO_2 stream produced by the Wellman-Lord Process into a more useful form. The SO_2 can be converted to sulfur by several processes that have been demonstrated or are under development. SO_2 can also be converted to sulfuric acid in a contact sulfuric acid plant. The acid production is less consumptive of fuel and reducing media, but the acid produced is more difficult to store and ship. The SO_2 Conversion Processes are discussed in detail in Section 3.8.

3.5 Magnesia Slurry Absorption Process

The Magnesia Slurry Absorption Process is a regenerable flue gas desulfurization process. SO_2 is removed from the flue gases by wet scrubbing with a slurry of magnesium oxide. Magnesium sulfite is the predominant species formed in the absorption reaction shown in Equation 3.5-1:

$$Mg(OH)_2 + SO_2 \rightarrow MgSO_3 + H_2O. \qquad (3.5\text{-}1)$$

The absorber effluent is centrifuged. The liquor is sent to the slurry tank for combination with makeup water, makeup MgO, and regenerated MgO to form the slurry feed for the scrubber. The magnesium sulfite cake is dried to remove free and bound water. Magnesium oxide is then regenerated in a calciner by thermal decomposition of the magnesium sulfite according to Equation 3.5-2:

$$MgSO_3 \underset{\Delta}{\rightarrow} MgO + SO_2. \qquad (3.5\text{-}2)$$

The concentrated SO_2 gas stream can be used to promote sulfuric acid or elemental sulfur.

3.5.1 Process Description

The design of the magnesia scrubbing system can be divided into five process areas:

1) Gas Pretreatment,
2) SO_2 Absorption,
3) $MgSO_3/MgSO_4$ Separation and Drying,
4) MgO Regeneration and SO_2 Recovery, and
5) Sulfur Production.

Figure 3.5-1 is a simplified flow diagram for the process.

Gas Pretreatment

Flue gas is pretreated in a venturi or tray-type prescrubber to cool and humidify the gas, and to reduce fly ash and chlorides. The humidification and cooling step prevents the evaporation of excessive amounts of water in the absorber. The

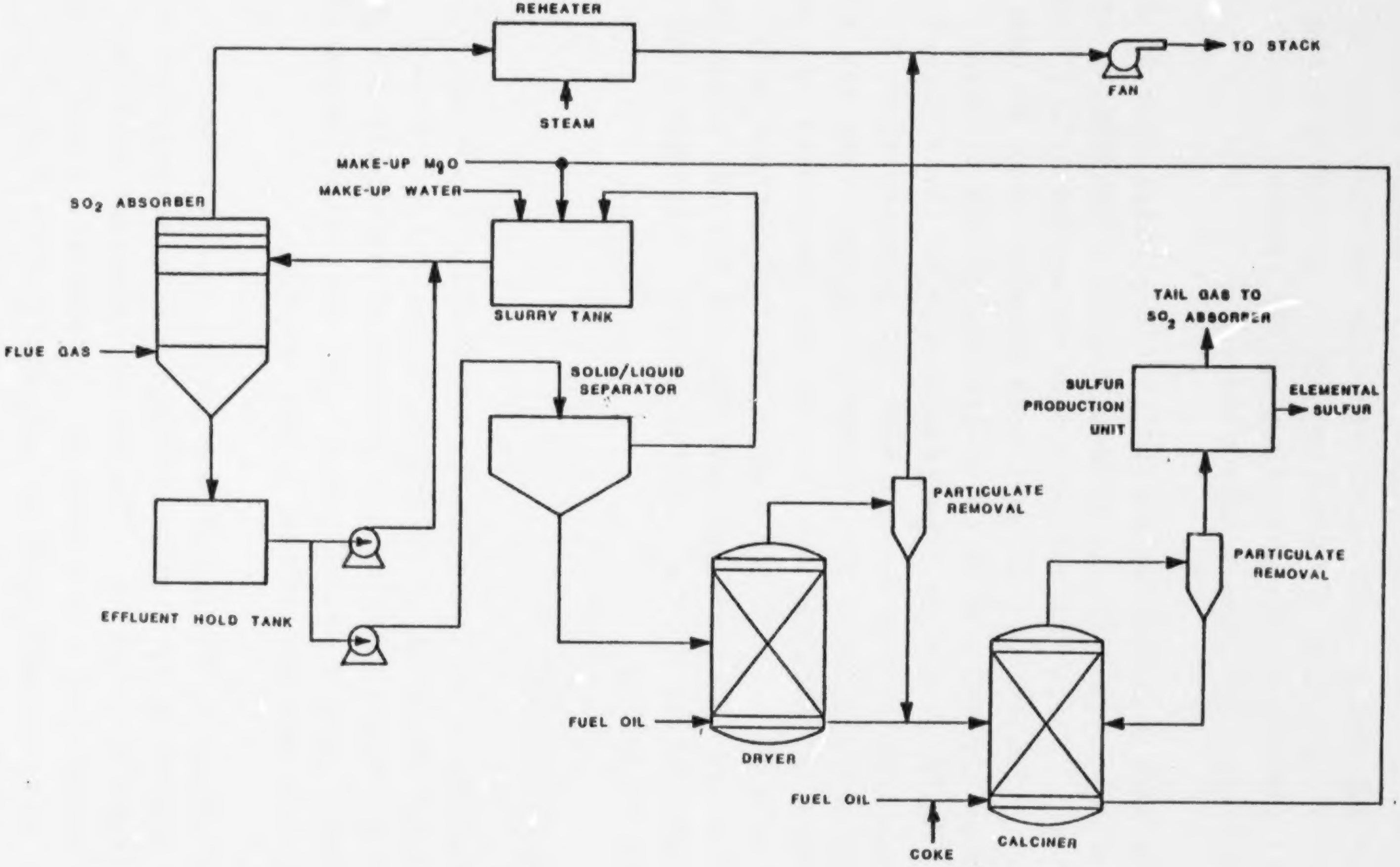
REHEATER
TO STACK
FAN
STEAM
MAKE-UP MgO
MAKE-UP WATER
SO2 ABSORBER
SLURRY TANK
FLUE GAS
SOLID/LIQUID SEPARATOR
TAIL GAS TO SO2 ABSORBER
SULFUR PRODUCTION UNIT
ELEMENTAL SULFUR
PARTICULATE REMOVAL
PARTICULATE REMOVAL
EFFLUENT HOLD TANK
FUEL OIL
DRYER
FUEL OIL
COKE
CALCINER

Figure 3.5-1. Process flow diagram for the magnesia slurry absorption process.

potential for scaling and plugging problems is reduced by the removal of fly ash which, containing vanadium and iron compounds, can catalyze the oxidation of $MgSO_3$ to $MgSO_4$. In a well designed prescrubber 99 percent of the chlorides can be removed, thus reducing the potential for stress corrosion

Flue gas, exiting the electrostatic precipitator at a temperature of 150°C (300°F), is passed through a venturi or tray-type prescrubber. There the gas is cooled to around 55°C (130°F) and humidified. The venturi is preferred because 70 to 80 percent of the remaining fly ash and 95 to 99 percent of the chlorides are removed. Although a tray-type prescrubber is satisfactory for cooling and humidifying the gas with low pressure drop, it provides lower reductions in fly ash and chlorides. The fly ash and other solids collected by the prescrubber are pumped to the ash disposal pond as a slurry of approximately 5 percent solids. Since absorption of chlorides and some SO_2 and SO_3 can cause the slurry to become fairly acidic, neutralization is accomplished with lime when necessary.

SO_2 Absorption

Development of the magnesia scrubbing process has followed at least three major technical routes since the early 1930's. Process variations include the use of magnesium sulfite/magnesium oxide slurries having a basic pH, the use of magnesium sulfite in acidic solution (clear liquor process).

The presence of manganese promotes desulfurization, oxidation of magnesium sulfite to sulfate, and decomposition of magnesium sulfate by roasting. The use of magnesium sulfite in an acidic solution (pH less than 6.0) produces a clear liquor which can be used on coal-fired systems where one scrubber is used to remove both the particulates and SO_2. A clear solution is used so that the ash can be filtered out. Because the vapor pressure of the scrubbing solution is higher, the process is restricted by a lower SO_2 removal efficiency than the basic slurry process. Since the basic slurry process is the most advanced system it will be evaluated in this study.

An aqueous slurry of magnesium hydroxide and magnesium sulfite (pH range 6.5 to 8.5) is used to absorb the SO_2 according to Equations 3.5-1 and 3.5-3:

$$Mg(OH)_2 + SO_2 \rightarrow MgSO_3 + H_2O, \qquad (3.5\text{-}1)$$

$$MgSO_3 + H_2O + SO_2 \rightarrow Mg(HSO_3)_2. \qquad (3.5\text{-}3)$$

Sulfite oxidation gives rise to sulfates in the system by the following reaction:

$$MgSO_3 + \tfrac{1}{2}O_2 \rightarrow MgSO_4. \qquad (3.5\text{-}4)$$

The test facility at Boston Edison reported sulfate concentration to be in the 15-20 weight percent range for solids shipped to the regeneration facility. As illustrated in the following equations, the sulfite and sulfate solids precipitate as hydrated crystals:

$$MgSO_3 + 6H_2O \rightarrow MgSO_3 \cdot 6H_2O, \qquad (3.5\text{-}5)$$

$$MgSO_3 + 3H_2O \rightarrow MgSO_3 \cdot 3H_2O, \qquad (3.5\text{-}6)$$

$$MgSO_4 + 7H_2O \rightarrow MgSO_4 \cdot 7H_2O. \qquad (3.5\text{-}7)$$

The bisulfite in the spent scrubbing liquor is reacted with magnesium hydroxide which is formed by slaking fresh and recycled magnesium oxide:

$$Mg(HSO_3)_2 + Mg(OH)_2 + 4H_2O \rightarrow 2(MgSO_3 \cdot 3H_2O), \qquad (3.5\text{-}8)$$

$$MgO + H_2O \rightarrow Mg(OH)_2. \qquad (3.5\text{-}9)$$

Cooled and humidified gas from the presrubber is passed through the absorption tower where the SO_2 level is reduced by at least 90 percent. The cleaned gas, reheated so that it has the proper dew point and buoyancy, is then exhausted to the atmosphere. The gas is reheated by heat exchange with high-pressure steam. Although alternatives to this reheating method exist, the use of steam allows coal to be used indirectly rather than premium fuels such as oil or natural gas.

When MgO slurry enters the absorption tower and absorbs SO_2, $MgSO_3$ crystals are formed. A bleed stream is sent to a centrifuge in the first step of MgO recovery. After makeup water, recovered MgO, and makeup MgO are added to the recycle slurry, the resulting slurry is returned to the absorber as scrubbing solution feed.

In a 500 MW plant, four 125 MW scrubber trains are used so that the scrubbers will be of reasonable size. This also enables the plant to run at a 375 MW capacity while maintenance is performed on one scrubber train.

$MgSO_3/MgSO_4$ Separation and Drying

A bleed stream is taken off the absorber effluent stream as a 10-15 percent solids slurry and sent to a thickener to produce a 40 percent solids slurry. This slurry is then centrifuged to recover a wet cake of $MgSO_3/MgSO_4$ hydrate crystals. The wet crystals are discharged from the centrifuge through a vertical chute into a screw feeder. The feeder acts as a seal and provides a continuous flow of wet solids into a rotary or fluid-bed dryer. The rotary kiln type dryer is presently used in the three U.S. magnesia scrubbing demonstration units. Combustion gases from an oil-fired burner are used to dry the crystals. A portion of the gases is recycled to the dryer chamber for temperature control; the remainder is exhausted to the stack. The dried $MgSO_3/MgSO_4$ is discharged from the dryer and conveyed to the calciner for MgO regeneration and SO_2 recovery.

MgO Regeneration and SO_2 Recovery

To generate MgO and SO_2, the dried $MgSO_3/MgSO_4$ crystals are calcined in an oil-fired rotary kiln or fluidized-bed reactor. The thermal decomposition reaction of $MgSO_3$ is shown in Equation 3.5-2:

$$MgSO_3 \underset{\Delta}{\rightarrow} MgO + SO_2. \qquad (3.5\text{-}2)$$

$MgSO_4$ is reduced in the presence of carbon as shown in Equation 3.5-10:

$$MgSO_4 + \tfrac{1}{2}C \underset{\Delta}{\rightarrow} MgO + SO_2 + \tfrac{1}{2}CO_2. \qquad (3.5\text{-}10)$$

Two installations have used rotary kilns to regenerate the magnesium oxide. However, rotary kilns have high dust losses and require a hot cyclone and venturi scrubber for magnesium solids recovery. In a fluidized-bed calciner, most of the MgO formed would go overhead with the SO_2 and combustion gases. Thus, this method would also require separation equipment.

The optimum calcining temperature in the reactor is set to achieve decomposition of all of the $MgSO_3/MgSO_4$ solids without "dead burning" the MgO. "Dead burned" MgO is chemically unreactive and not effective for further SO_2 removal. Operating temperatures in the 800°C (1500°F) range have been used in the rotary calciner.

SO_2 Conversion

After dust removal, the sulfur dioxide-rich gas from the calciner is sent to either a sulfur or sulfuric acid production unit. The gas stream from the magnesium oxide calciner is well suited for sulfuric acid production. The calciner off-gas is saturated with water at 40°C (100°F) and contains about 8-10 percent SO_2. Acid production is less consumptive of fuel and reducing media, but the acid produced is more difficult to store and ship. The SO_2 can also be converted to elemental sulfur by several processes that have been demonstrated or are under development. The SO_2 Conversion Processes are discussed in detail in Section 3.8.

3.6 Double Alkali Wet Scrubbing

The double alkali flue gas desulfurization process is a "throwaway" process that removes SO_2 from the flue gas by wet scrubbing with a sodium sulfite liquor. In a second step, a waste sludge of calcium sulfite and sulfate is formed, as is a regenerated sodium sulfite scrubbing liquor. Separating the absorption and waste production steps has the advantage of scrubbing the flue gas with a more concentrated soluble alkali. This permits the use of lower liquid-to-gas ratios in the double alkali process as compared with lime/limestone processes. Furthermore, the amount of soluble and slurried calcium in the scrubber is minimized, thus offering the opportunity for better scale control.

3.6.1 Process Description

A number of processes can be considered double alkali processes. In the United States, most of the developmental work has emphasized sodium-based double alkali systems using lime for regeneration. Double alkali systems using an ammonia/calcium base have been tested, but they suffer the disadvantage of potentially producing a visible ammonium salt plume from the scrubbing system. The following process description will be limited to sodium/calcium-based processes.

The design of a double alkali system can be divided into four process areas:

1) SO_2 Absorption,
2) Waste Production and Sorbent Regeneration,
3) Solids Separation, and
4) Solids Disposal.

Figure 3.6-1 shows a generalized flow diagram for a double alkali system.

SO_2 Absorption

Normally, gas from the electrostatic precipitator passes through an absorption tower, where SO_2 is removed by absorption into a sodium hydroxide or sodium sulfite scrubbing solution. The gas may be saturated in a presaturation section of the absorber before it enters the absorber itself. The cleaned gas, reheated to 80°C (175°F) so that it has the proper dew point and buoyancy, is then exhausted to the atmosphere. The scrubber effluent liquor is regenerated with lime or

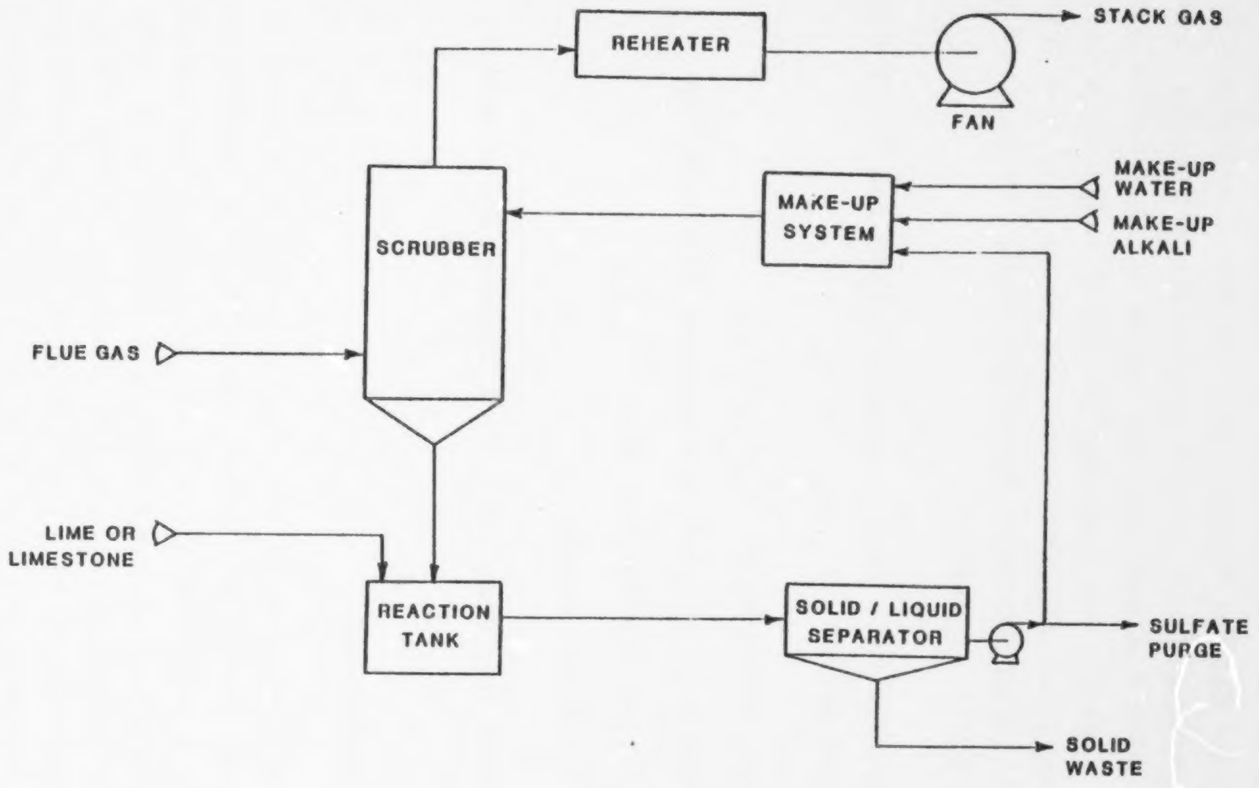
REHEATER
STACK GAS
FAN
MAKE-UP WATER
MAKE-UP ALKALI
MAKE-UP SYSTEM
SCRUBBER
FLUE GAS
LIME OR LIMESTONE
REACTION TANK
SOLID / LIQUID SEPARATOR
SULFATE PURGE
SOLID WASTE

Figure 3.6-1. Process flow diagram for the double-alkali scrubbing process.

limestone in a reaction tank. The calcium sulfite and sulfate solids formed in the reaction tank are removed from the system in a solid/liquid separator. The separator liquor is recycled to the absorber.

A prescrubber (for gas saturation and chloride removal) on a separate liquor loop is not normally used in a double alkali system as it is in a Wellman-Lord or Magnesia Slurry system because of water balance problems. If a prescrubber is used, the major water loss in the system (evaporation by the flue gas) occurs in the prescrubber. The only water loss in the remainder of the system is the water lost with the solid waste. This small loss may not allow addition of enough water for lime slaking, Na_2CO_3 solution, and cake washing. The limitation of cake washing would result in a high sodium make-up requirement and a high dissolved solids content in the waste. Normally, enough chloride is removed in the liquor discharged with the solid waste to prevent excessive chloride build-up. In applications with coals having a very high chloride content, however, this mechanism for chloride removal may not be adequate, and a prescrubber may be required for chloride removal in spite of the previously mentioned difficulties.

The principal reactions for the absorption of SO_2 from the flue gas are shown in equations 3.6-1 and 3.6-2:

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O, \qquad (3.6\text{-}1)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3. \qquad (3.6\text{-}2)$$

In double alkali systems where lime is used as the regenerant, the pH range varies over the hydroxide/sulfite/bisulfite range so that reactions indicated by Equations 3.6-1 and 3.6-2 both occur. In double alkali systems using limestone as the regenerant, the pH range is limited to the sulfite/bisulfite range and only the second reaction (Equation 3.6-2) occurs.

A very important side reaction is the oxidation of sulfite to sulfate caused by the absorption of oxygen in the flue gas:

$$Na_2SO_3 + \tfrac{1}{2}O_2 \rightarrow Na_2SO_4. \qquad (3.6\text{-}3)$$

The absorber effluent is sent to the reaction tank.

The double alkali processes can be operated in either the "dilute" or "concentrated" mode. Here these terms refer to the concentration of active alkali (sulfite). In general, dilute systems are more suited to applications in which oxidation is expected to be relatively high, whereas concentrated systems are favored in applications where oxidation is expected to be low.

In order to minimize the potential for gypsum ($CaSO_4 \cdot 2H_2O$) scaling in the scrubber, a softening step is used to reduce the dissolved calcium concentration in the scrubber feed liquor. In dilute double alkali systems, carbonate softening is generally employed. Such systems utilize soda ash (Na_2CO_3) and carbon dioxide (CO_2) to precipitate the dissolved calcium as carbonate:

$$Na_2CO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaCO_3, \qquad (3.6\text{-}4)$$

$$CO_2 + Ca(OH)_2 \rightarrow CaCO_3 + H_2O. \qquad (3.6\text{-}5)$$

The main function of the soda ash is to replenish the sodium losses in the system (solid disposal). The carbon dioxide supplies the additional softening required. In concentrated systems, sulfite softening is used:

$$Na_2SO_3 + Ca(OH)_2 + \tfrac{1}{2}H_2O \rightarrow 2NaOH + CaSO_3 \cdot \tfrac{1}{2}H_2O. \qquad (3.6\text{-}6)$$

From economic and chemical utilization standpoints, the system should be operated with minimum softening to avoid scrubber scaling.

Waste Production and Sorbent Regeneration

The scrubber effluent liquor is sent to a reaction tank where the sorbent is regenerated with lime or limestone. Solid calcium sulfite and sulfate are formed as shown for the lime system by equations 3.6-7 through 3.6-9:

$$Ca(OH)_2 + 2NaHSO_3 \rightarrow Na_2SO_3 + CaSO_3 \cdot \tfrac{1}{2}H_2O + \tfrac{3}{2}H_2O, \qquad (3.6\text{-}7)$$

$$Ca(OH)_2 + Na_2SO_3 + \tfrac{1}{2}H_2O \rightarrow 2NaOH + CaSO_3 \cdot \tfrac{1}{2}H_2O, \qquad (3.6\text{-}8)$$

$$Ca(OH)_2 + Na_2SO_4 + 2H_2O \rightarrow 2NaOH + CaSO_4 \cdot 2H_2O. \qquad (3.6\text{-}9)$$

The corresponding reactions for the limestone system are given by equations 2.6-10 and 3.6-11:

$$CaCO_3 + 2NaHSO_3 \rightarrow Na_2SO_3 + CaSO_3 \cdot \tfrac{1}{2}H_2O + CO_2 + \tfrac{1}{2}H_2O, \qquad (3.6\text{-}10)$$

$$CaCO_3 + 2NaHSO_4 + H_2O \rightarrow Na_2SO_4 + CaSO_4 \cdot 2H_2O + CO_2 \quad (3.6\text{-}11)$$

Solids Separation

The calcium sulfite and calcium sulfate solids formed in the reaction tank are withdrawn from the system in a solid/liquid separator. After make-up alkali and water are added, the separator effluent liquor is recycled to the scrubbing loop. A liquid purge stream is required to remove soluble sodium sulfate. Failure to allow for sulfate removal from double alkali systems will ultimately result in 1) precipitation of sodium sulfate somewhere in the system if active sodium is made up to the system or 2) in the absence of make-up, eventual deterioration of the SO_2 removal capability due to the loss of active sodium from the system.

Solids Disposal

The double alkali flue gas desulfurization process is a non-regenerative or "throwaway" process. Sludge disposal is one of the main disadvantages of "throwaway" FGD systems when compared to "recovery" processes. The quantity of sludge produced is large in weight and volume, and requires a large waste pond or landfill area for disposal.

On-site disposal is usually performed by sending the waste solids to a large pond. Settling of the solids occurs and pond water is recycled to the process hold tank for reuse.

Stabilization methods are currently under development to convert the sludge to structurally stable, leach-resistant, landfill material. These methods could be used when on-site disposal is not possible. The stablized material can then be trucked to an off-site area for landfill.

3.7 Physical Coal Cleaning

First practiced in the United States in 1880, coal cleaning is an established technology in the mining industry. Since 1945 the annual percentage of coal cleaned has tripled. At present, nearly 50 percent of the annual U.S. coal production is physically cleaned. To date, coal cleaning operations have not been utilized to control sulfur oxide emissions. The primary functions of coal preparation plants have been to remove rock and ash from coal and to produce coking grade coals for use in metallurgical processes. In these operations, sulfur removal has not been optimized. However, pilot plant studies to optimize sulfur removal have been reported; currently General Public Utilities Corporation and New York Electric and Gas Corporation are constructing a full-scale facility to achieve compliance with sulfur oxide emission regulations (DA-189).

Physical coal cleaning removes impurities from coal via a mechanical separation process. In most cleaning operations, this separation of impurities is based on a gravity difference between coal (which is relatively light) and contaminants such as pyrite (FeS_2), ash, and rock (which are heavier) (PA-003).

Sulfur occurs in a coal seam in three forms: pyritic, organic, and sulfate. In any given coal the amount of sulfate sulfur is negligible. The total sulfur content may vary from less than 1 percent to over 8 percent, with most coals in the 2 to 5 percent range. The total sulfur content distribution between the organic and pyritic forms ranges from 5 to 60 percent and 40 to 95 percent, respectively.

Since organic sulfur in coal is chemically bound and requires a chemical extraction process for removal, physical coal cleaning is restricted to removal of the pyritic sulfur from coal.

The pyritic sulfur content of coal is present in many forms and particle sizes. These range from coarse sized particles, which are relatively easy to remove, to finely disseminated particles which could almost be classified as inherent. Not only is the distribution and nature of pyrite particles different in various coal seams, but the amount and variety of sulfur forms will vary in the same coal bed from one area to another. Therefore, it is difficult to draw conclusions as to which physical cleaning methods will best suit a particular coal seam.

The potential for sulfur reduction in coal by application of conventional physical coal cleaning technology is limited. The U.S. Bureau of Mines investigated the sulfur release potential of 322 coal mines representing a large percentage of utility coal sources. The results of this study indicated that the physical removal of sulfur is both coal and process specific (DE-064). Individual coals respond uniquely to the various unit operations utilized in cleaning facilities.

The generalized results obtained from the 322 mines studied in the Bureau of Mines investigation indicate that an average cleaning process operating at 90 percent yield has the potential to reduce the total sulfur content of the coal by 30 percent. Of the 322 mines sampled, less than 30 percent of the coal could be cleaned to a sulfur content of 1 percent (DE-064).

3.7.1 Process Description

Although it is not possible to describe a universally applicable coal cleaning process, certain processing areas which are common to most cleaning operations can be identified. Figure 3.7-1 is a flow diagram of a coal cleaning facility depicting common process areas without detailing specific unit operations (LO-071, CO-380).

The following process areas are found in most coal cleaning facilities. Listed under each area are various operations which may be utilized in an individual cleaning process.

Initial Coal Preparation

1) Storage
    - Bins, Silos, and Hoppers

2) Rough Cleaning/Primary Breaking
    - Tramp Iron Removal
    - Scalping Screens
    - Crushing Equipment

3) Raw Coal Sizing
    - Shaking Screens
    - Vibrating Screens

Fine Coal Processing

1) Dry Cleaning
    - Airflow Cleaner
    - Air Tables
    - Centrifugal/Electrostatic Separator

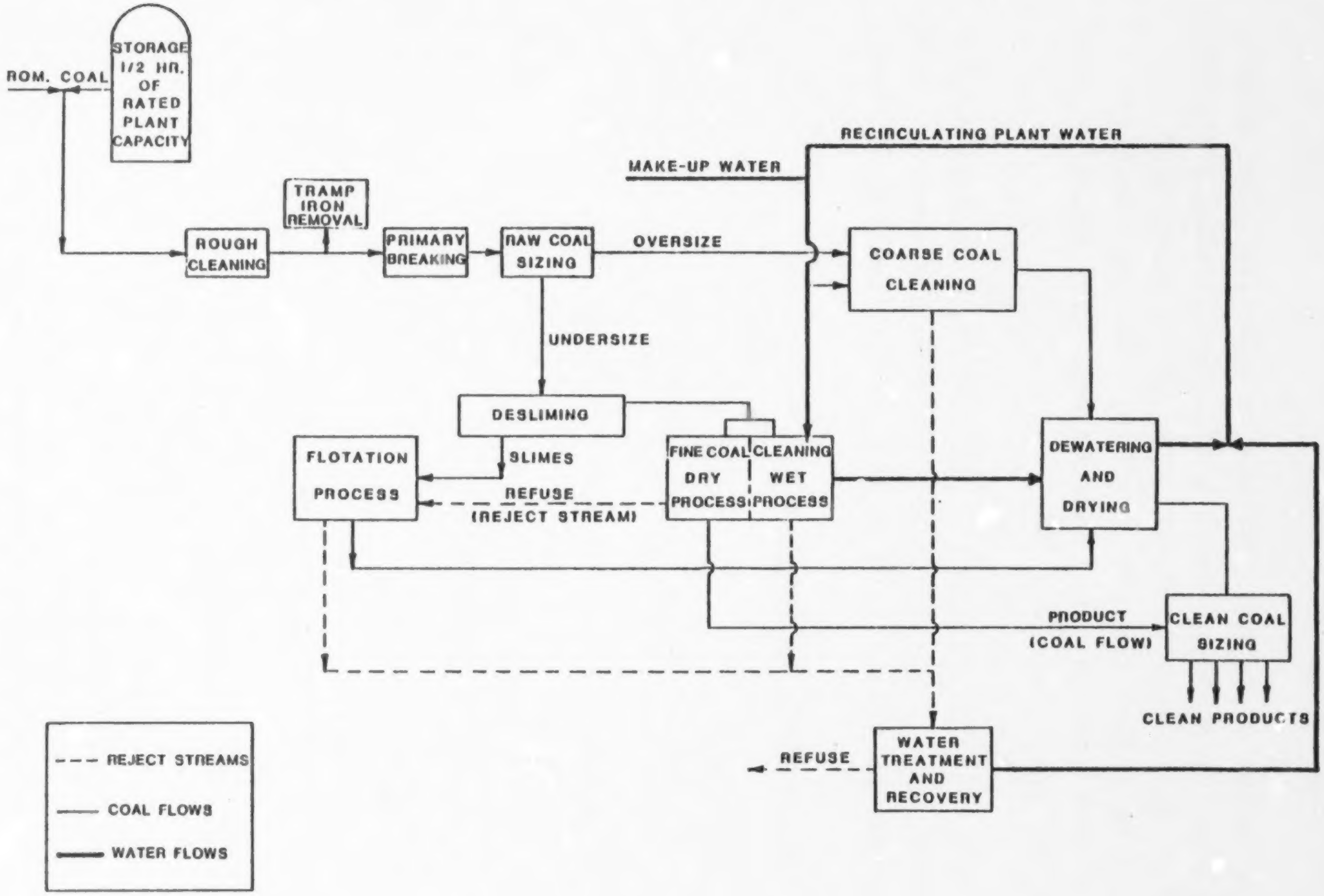
ROM. COAL
STORAGE 1/2 HR. OF RATED PLANT CAPACITY
ROUGH CLEANING
TRAMP IRON REMOVAL
PRIMARY BREAKING
RAW COAL SIZING
OVERSIZE
UNDERSIZE
MAKE-UP WATER
RECIRCULATING PLANT WATER
COARSE COAL CLEANING
DESLIMING
SLIMES
FLOTATION PROCESS
REFUSE (REJECT STREAM)
FINE COAL CLEANING
DRY PROCESS
WET PROCESS
DEWATERING AND DRYING
PRODUCT (COAL FLOW)
CLEAN COAL SIZING
CLEAN PRODUCTS
WATER TREATMENT AND RECOVERY
REFUSE
REJECT STREAMS
COAL FLOWS
WATER FLOWS

Figure 3.7-1. Generalized coal cleaning process.

2) Wet Cleaning
    - Jigs
    - Heavy-Medium Cyclones
    - Water Tables
    - Hydrocyclones
    - Spiral Classifiers
    - Froth Flotation

3) Desliming

Coarse Coal Processing

- Jigs
- Heavy-Media Separators
- Hydrocyclones
- Launders

Water Management/Refuse Disposal

1) Dewatering
    - Sieve Bends
    - Screens
    - Thickners
    - Cyclones

2) Drying
    - Centrifuges
    - Filters
    - Thermal Dryers

3) Water Recovery

4) Refuse Disposal
    - Fire Prevention
    - Prevention of Ground Water Pollution

1) Initial Coal Preparation

Prior to the actual cleaning process, run-of-mine (R.O.M.) coal must undergo initial preparation. This involves preliminary crushing of the coal to remove large rock fractions and to liberate entrained impurities such as clay, rock, and other inorganic materials, including pyrite. The first crushing step is followed by a screening operation and secondary crushing. A second screening step produces two product streams from this process area: one containing a fine fraction (usually less than 6.5 mm) and the other containing coarse particles (nominally 76 x 6.5 mm). These two coal flows are then fed to their respective process areas where the actual cleaning operation takes place (CO-380, LO-071).

2) Fine Coal Processing

The fine coal processing area of a preparation plant can employ both wet and dry cleaning operations. In plants utilizing a dry coal cleaning process, fine coal from the initial preparation step flows to a feed hopper and then to an air cleaning operation. This cleaning operation can employ one of several devices which rely on an upward current of air traveling through a fluidized bed of crushed coal. Separation is effected by particle size and density. Product streams from a dry cleaning process are sent directly to the final coal preparation step, while reject streams are usually processed further in wet cleaning operations (FI-102).

In operations utilizing wet methods to effect fine coal cleaning, the process feed stream of less than 6.5 mm coal is slurried with water as it enters the fine coal processing area of the plant. This slurry is then subjected to a desliming operation which removes a suspension containing approximately 50 percent of minus 200 mesh material (FI-102). The cutoff size for this separation is usually in the range of 28 to 48 mesh (PE-030). This desliming operation is necessary because the presence of slimes adversely affects the capacity and efficiency of fine cleaning units.

Subsequent to desliming, the oversize coal fraction (greater than 28 mesh) is pumped to the fine coal cleaning process. Here, fine coal particles undergo gravity separation in one of several wet cleaning devices. This removes a percentage of ash and pyritic sulfur to produce a clean coal product. The product stream from this operation is fed to the drying area of the plant; refuse material is further processed in the water treatment section.

The slimes removed from the fine coal stream are fed to a froth flotation process. Other material, such as reject from dry cleaning operations, may also be treated in the flotation process. This process consists of "rougher" and "cleaner" sections which are comprised of cells of flotation machines. Upon entering the flotation process area, the slime suspension is treated with a frothing agent. This agent selectively floats coal particles in the flotation machines while allowing pyrite and ash impurities to settle. Processing slime in the "rougher" cells produces a reject stream and a low grade product. The low grade product is further processed in the "cleaner" cells to produce a clean coal product. This final float product is then sent to the dewatering area for further handling, while reject material from both rougher and cleaner sections is processed in the water treatment and recovery area.

3) Coarse Coal Processing

Feed to the coarse coal processing area of the plant consists of oversize material (76 x 6.5 mm particles) from the initial preparation area. This feed stream is slurried with water prior to cleaning, since coarse coal cleaning operations employ wet processing equipment to remove impurities from the coal. The coarse coal slurry is fed to one of the many types of process equipment currently employed in coarse coal cleaning. Here, impurities are separated from the coal due to differences in product and reject density. It is common practice to remove a middling fraction from the separation operation and process it further by means of recycle or by feed to another cleaning proces. These cleaning operations result in removal of two streams from the coarse coal processing area: a product and reject stream. Subsequent to the coarse cleaning operation, the product stream is pumped to the dewatering and drying area of the plant, while the reject stream is processed in the water treatment and recovery area.

4) Water Management/Refuse Disposal

Dewatering and drying equipment handle the product flows from both the fine and coarse coal preparation areas. Typically, cleaning plants employ mechanical dewatering operations to separate coal slurries into a low-moisture solid and a clarified supernatant. The solid coal sludge produced in the dewatering step can be mechanically or thermally dried to further reduce the moisture. The supernatant from the dewatering process is returned to the plant's water circulation system.

The water treatment and recovery section of a cleaning plant processes refuse slurries containing both coarse material and reject slimes. Here, the refuse slurry is dewatered, typically in thickeners and settling ponds. The supernatant from this operation is returned for reuse in the plant, while the refuse is buried and revegetated to prevent burning.

The coal product from the dewatering and drying area of the plant can be further processed. This may involve crushing and screening operations to separate the product into various product sizes. The cleaned and sized product is then loaded into rail cars for shipping.

3.8 SO_2 Conversion Processes

The Wellman-Lord and the Magnesium Slurry Wet Scrubbing Processes produce an SO_2 rich product gas stream which can be converted to either sulfuric acid or elemental sulfur. Several available processes and several processes in the developmental stages are suitable for FGD by-product production. In a recent study for EPRI (OT-051), Radian chose the Allied Chemical Sulfur Plant, and a single absorption contact sulfuric acid for economic comparisons on the basis of data availability. These two processes will be described, and the subsequent water effluent and water consumption quantities evaluated.

3.8.1 Sulfuric Acid Production

On a capital cost basis, a single absorption contact process is the suggested method for conversion of FGD SO_2 streams into sulfuric acid. The tail gas from the acid unit would be sent back to the SO_2 scrubber to prevent SO_2 emissions. An alternative to treating the tail gas would be the use of a

Double Contact/Double Absorption Acid Unit (DC/DA). This facility would result in added costs to the acid production unit whereas routing the tail gas back through the FGD process would result in increased FGD costs for processing the extra gas. An economic comparison on a site specific basis is needed to determine the best choice. In this study, it is assumed that the FGD system can handle the relatively small increase (about 5 percent) in load from routing the gases back through the system.

Process Description

Basically, the single absorption contact process oxidizes sulfur dioxide to sulfur trioxide over a vanadium pentoxide catalyst. The sulfur trioxide then combines with water in an absorber where the product acid is formed. Single absorption contact acid plants operate at sulfur dioxide conversion efficiencies of about 97 percent. A typical flow diagram of the process is illustrated in Figure 3.8-1.

The feed gas to the process should consist of 8.4 to 9.0 volume percent SO_2 and 8.6 to 9.2 volume percent O_2 on a dry basis. This feed gas first passes through a drying tower. The dry gas then passes through a series of heat exchangers to bring the gas temperature to 435°C (815°F). The hot gas passes through the first three catalyst beds with intermediate heat exchange to remove the heat generated by the exothermic reaction shown below:

$$SO_2 + \tfrac{1}{2}O_2 \xrightarrow{\text{catalyst}} SO_3. \qquad (3.8\text{-}1)$$

After the third catalyst bed, the gas is fed to an absorber where the SO_3 combines with the water in the circulating acid. Before being mixed with the flue gas going to the

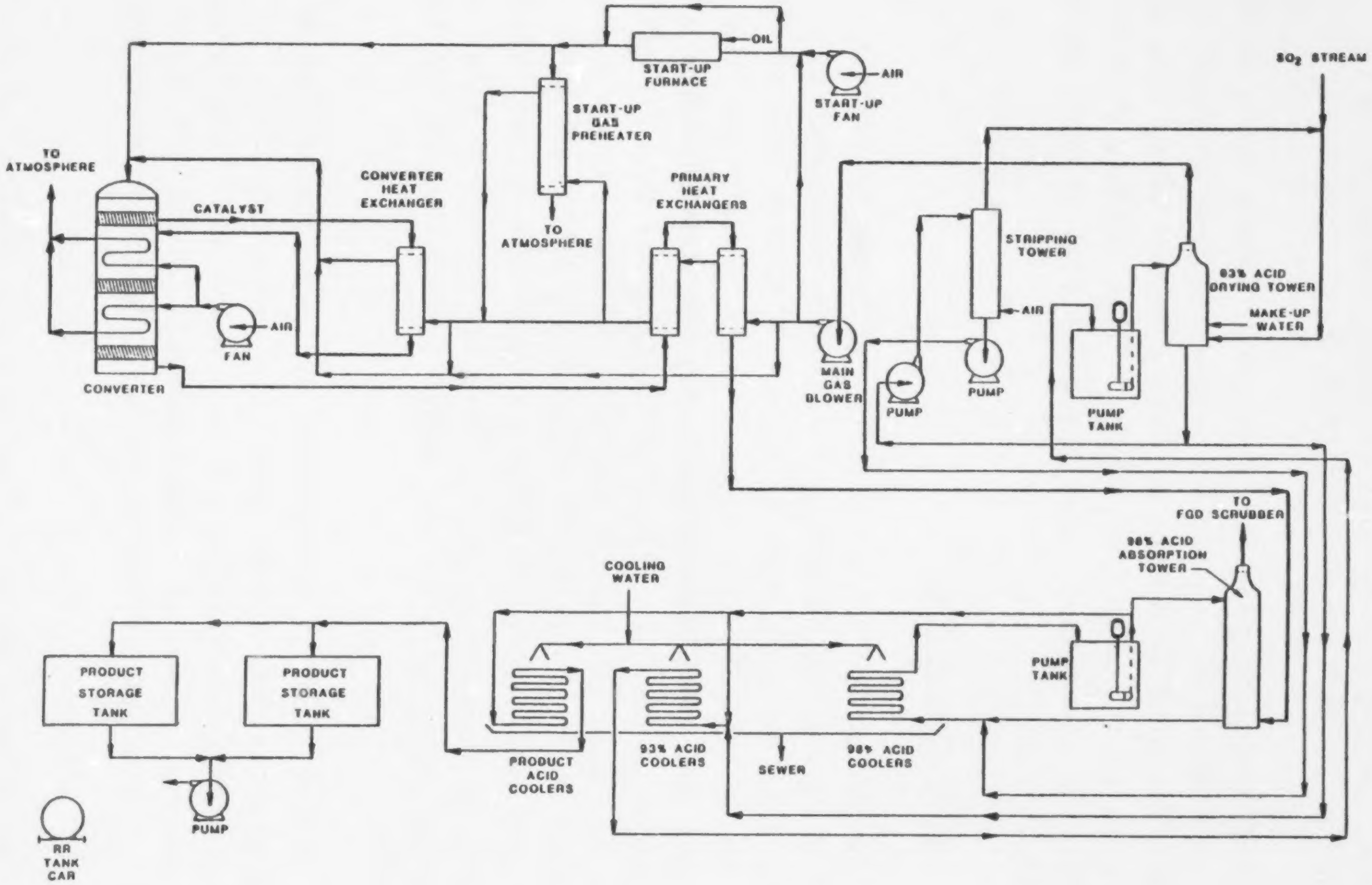
TO ATMOSPHERE
CATALYST
CONVERTER HEAT EXCHANGER
START-UP FURNACE
OIL
START-UP GAS PREHEATER
TO ATMOSPHERE
PRIMARY HEAT EXCHANGERS
AIR
START-UP FAN
SO_2 STREAM
STRIPPING TOWER
93% ACID DRYING TOWER
MAKE-UP WATER
AIR
FAN
AIR
CONVERTER
MAIN GAS BLOWER
PUMP
PUMP
PUMP TANK
TO FGD SCRUBBER
98% ACID ABSORPTION TOWER
COOLING WATER
PRODUCT STORAGE TANK
PRODUCT STORAGE TANK
PUMP TANK
PRODUCT ACID COOLERS
93% ACID COOLERS
SEWER
98% ACID COOLERS
PUMP
RR TANK CAR

Figure 3.8-1. Typical flow diagram of a single absorption contact sulfuric acid plant.

SO_2 absorber, the vapor from this absorber passes through a demister to remove any entrained acid mist.

Sulfuric acid from the absorption towers flows to the acid pump tank where it is diluted to the required strength for absorption. The acid is pumped through a series of coolers before returning to the absorption towers. The product acid is removed from the absorption tower system through product coolers and sent to storage.

3.8.2 Sulfur Production

Elemental sulfur is currently produced by reacting an SO_2 stream with a reducing agent at elevated temperatures. Three processes have been applied to an SO_2 source from an FGD process: the Allied Chemical Process using natural gas, the RESOX Process using anthracite coal, and the BAMAG Process using a medium Btu town gas. In addition, preliminary work is being done to investigate the use of coal gasification reducing gas in the Allied Chemical Process. Because the RESOX and BAMAG processes are in early developmental stages, the Allied Chemical Process will be selected as the process for elemental sulfur production for water effluent and water consumption evaluations in our study.

Allied Chemical Corporation has developed and commercialized a process for direct, catalytic reduction of SO_2 to elemental sulfur using natural gas as a reductant. The first plant to use the process operated successfully for two years using a 12 percent SO_2 stream from a sulfide-ore roasting facility.

The process may be joined to a regenerable FGD process that produces a SO_2-rich gas stream with a low oxygen content.

This application is being demonstrated at the D. H. Mitchell Station of Northern Indiana Public Service Co. (NIPSCO) at Gary, Indiana. There, the process has been combined with the Wellman-Lord SO_2 Recovery Process to provide an FGD system for a 115 MW coal-fired boiler. Acceptance testing is scheduled for July 1977.

Process Description

A process flow diagram for an Allied Chemical Process with an SO_2 feed stream from an FGD system is shown in Figure 3.8-2. The plant consists of three main sections: 1) gas purification, 2) SO_2 reduction, and 3) sulfur recovery. The gas purification system, which is designed to remove excess water vapor and gaseous and solid impurities, is not required for all of the FGD processes.

The principle function of the catalytic reduction section is to increase the H_2S/SO_2 ratio in the gas stream to approximately the stiochiometric ratio of 2:1 required for the Claus reaction, while achieving maximum formation of elemental sulfur. The primary reaction system may be summarized in the following equations:

$$CH_4 + 2SO_2 \rightarrow CO_2 + 2H_2O + S_2 \qquad (3.8\text{-}2)$$

$$4CH_4 + 6SO_2 \rightarrow 4CO_2 + 4H_2O + 4H_2S + S_2 \qquad (3.8\text{-}3)$$

In the reduction section, the SO_2 stream, which has been combined with preheated natural gas, first passes through a four-way flow reversing valve and a final preheat reactor. The heated stream then enters the primary reactor where over 40 percent of the total recovered sulfur is formed. The reactor, which uses a

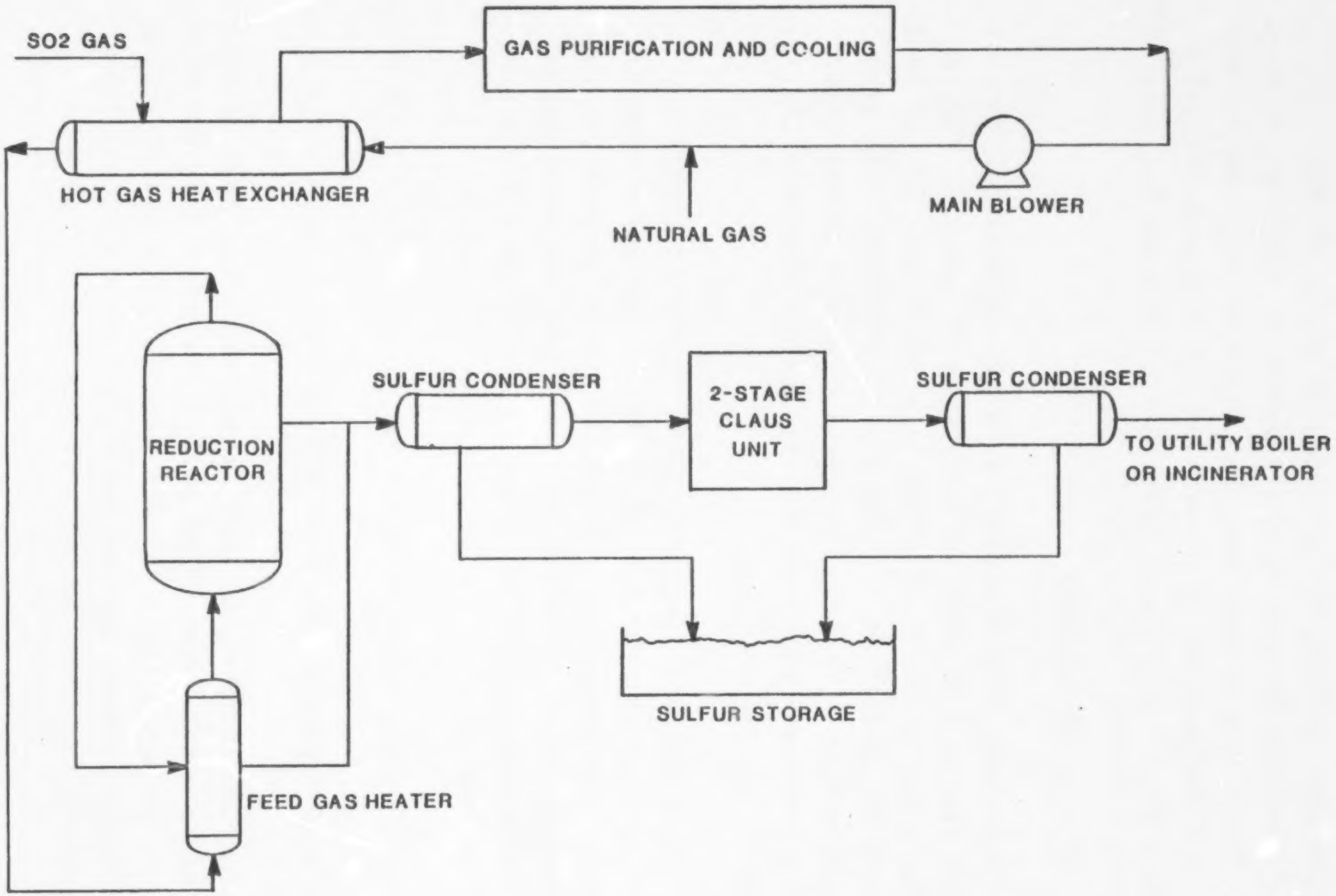
SO2 GAS
GAS PURIFICATION AND COOLING
HOT GAS HEAT EXCHANGER
NATURAL GAS
MAIN BLOWER
SULFUR CONDENSER
2-STAGE CLAUS UNIT
SULFUR CONDENSER
REDUCTION REACTOR
TO UTILITY BOILER OR INCINERATOR
SULFUR STORAGE
FEED GAS HEATER

Figure 3.8-2. Typical process flow diagram of the allied chemical SO_2 reduction process.

catalyst developed by Allied Chemical that is stable up to 1100°C (2000°F), achieves efficient methane utilization and provides minimum formation of undesirable side products. Careful control of the reaction conditions is necessary to achieve chemical equilibrium in the single reactor.

From the reduction reactor, the gas passes through a second heat regenerator where it gives up its heat to the packing. Direction of flow through the two heat regenerators is periodically reversed to interchange their functions of heating and cooling.

After condensing sulfur in a steam generator, the gas stream enters a two-stage Claus reactor system where H_2S and SO_2 react to produce elemental sulfur and water. At this point, product sulfur is again removed from the gas by condensation. Residual H_2S in the Claus plant effluent gas is oxidized to SO_2 by recycling the gas stream back to the boiler. The residual SO_2 is then recovered in the absorber of the original FGD recovery process.

According to the developer, this process can be applied directly to SO_2 streams containing as low as four percent SO_2, where the oxygen content is not over five percent. Processing streams with low SO_2 concentrations will, however, be costly as compared to more concentrated streams. When higher oxygen concentrations are encountered, provisions must be made to dissipate the excess heat produced as a result of methane oxidation. The Canadian plant has demonstrated that this process is capable of converting better than 90 percent of the SO_2 from the entering gas stream. Operation at one-third of design capacity with constant operation efficiency has been established. The major disadvantage of this method of sulfur production is the need for methane as a reductant.

Since the availability of methane in the future is predicted to decrease substantially, a process using a CO/H_2 coal gasification reducing gas appears to be an attractive alternate route to elemental sulfur production. Allied Chemical has made a preliminary evaluation of using a CO/H_2 reducing gas. The primary reactions in such a process would be as follows:

$$2CO + SO_2 \rightarrow 2CO_2 + \tfrac{1}{2}S_2 \qquad (3.8\text{-}4)$$

$$2H_2 + SO_2 \rightarrow 2H_2O + \tfrac{1}{2}S_2 \qquad (3.8\text{-}5)$$

The configuration of such a process would probably be very similar to the present Allied Chemical process using methane reductant.

4.0 WATER CONSUMPTION

This section assesses the impact of SO_x control technologies on coal-fired power plant water consumption. Many of the systems in a coal-fired power plant, the FGD systems under study, the physical coal cleaning process, and the SO_2 conversion processes, have large circulating water requirements. Fresh water makeup is required due to evaporation, solids occlusion, and blowdown losses. This section presents the results of calculations to determine the effect of the SO_x control technologies on raw water makeup requirements. First, the water consumption of a power plant uncontrolled for SO_x emissions is calculated. Then, the water consumption for each of the various SO_x control strategies is calculated. A matrix presentation of the results by model plant system is made. A more detailed presentation of the calculations is made in Appendix A. A base case 500 MW power plant is used for discussion.

4.1 Coal-Fired Power Plant Water Consumption; Uncontrolled for SO_x Emissions

The systems and operations in a coal-fired power plant that require fresh water makeup are the

- cooling water system,
- ash handling system,
- boiler makeup,
- water conditioning operations,
- equipment cleaning operations, and
- miscellaneous.

Due to the general unavailability of data on intermittent and miscellaneous water consumptive practices, water conditioning, equipment cleaning, and miscellaneous operations will be addressed as "general service water". Plant data from water recycle/reuse studies conducted by Radian for the EPA (NO-106, NO-137, HA-636, GA-203, CH-387) will be used to characterize these requirements.

Current water management in the power generation industry uses two basic processes: 1) once-through, and 2) recirculatory. Plant layout and water management practices vary widely in power plant water systems. Two extremes in water management are those in which: 1) once-through techniques are used in all systems, and 2) recirculatory practices are maximized in all systems. The total recirculatory system ultimately results in zero-discharge. The once-through system requires/discharges enormous quantities of water. Because of costs associated with water conditioning operations and wastewater treating, some recirculatory use of water is common practice. And because of the potential for scaling and fouling of lines, total recirculatory use of water is limited when fresh water is readily available. As the nation approaches a goal of zero discharge, recirculatory systems are becoming more prevalent. To characterize the range of water consumption patterns typical of current practices, Radian has chosen to characterize four model power plant water management systems. These represent the full range of power plant water management systems used today. The four systems are:

System #1: All power plant water systems are once-through; Figure 4.1-1

System #2: Recirculatory cooling at 2.5 cycles of concentration, once-through ash handling, and once-through general services water; Figure 4.1-2.

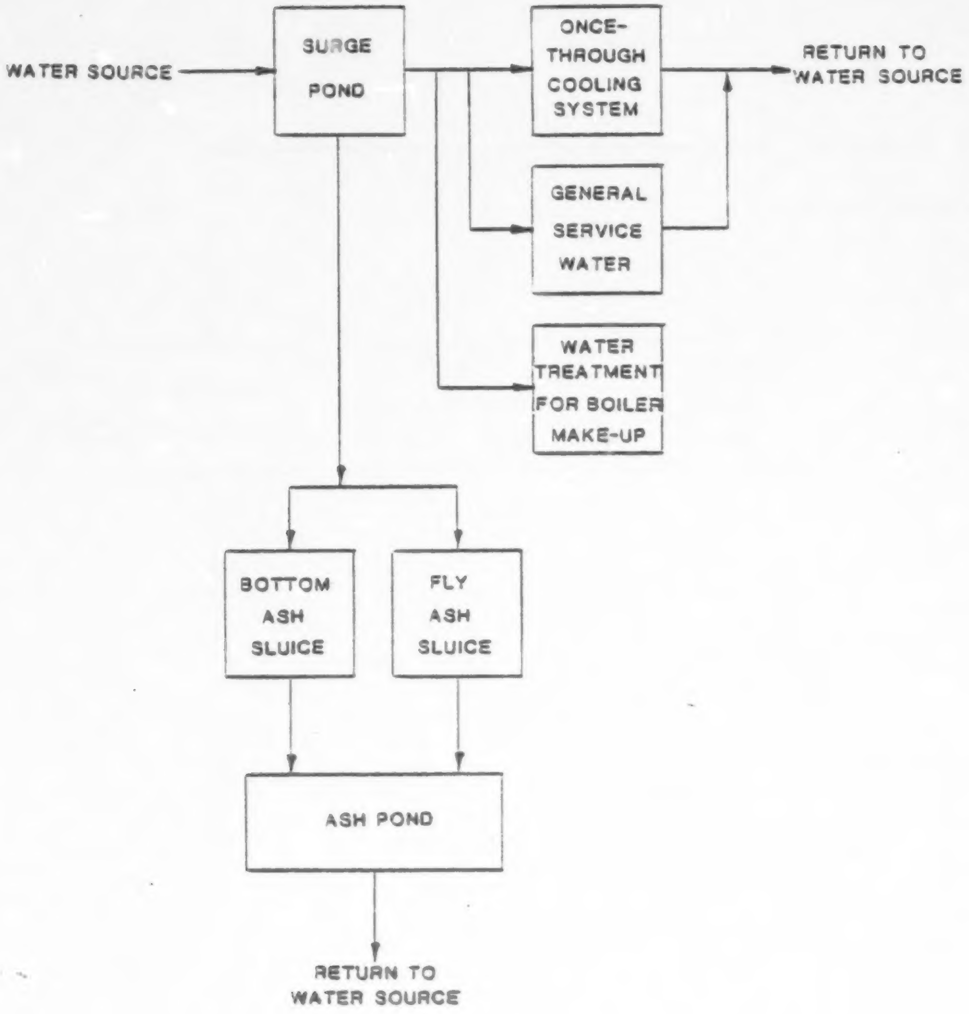
WATER SOURCE
SURGE POND
ONCE-THROUGH COOLING SYSTEM
RETURN TO WATER SOURCE
GENERAL SERVICE WATER
WATER TREATMENT FOR BOILER MAKE-UP
BOTTOM ASH SLUICE
FLY ASH SLUICE
ASH POND
RETURN TO WATER SOURCE

Figure 4.1-1. Power plant water system: system #1 once-through water management.

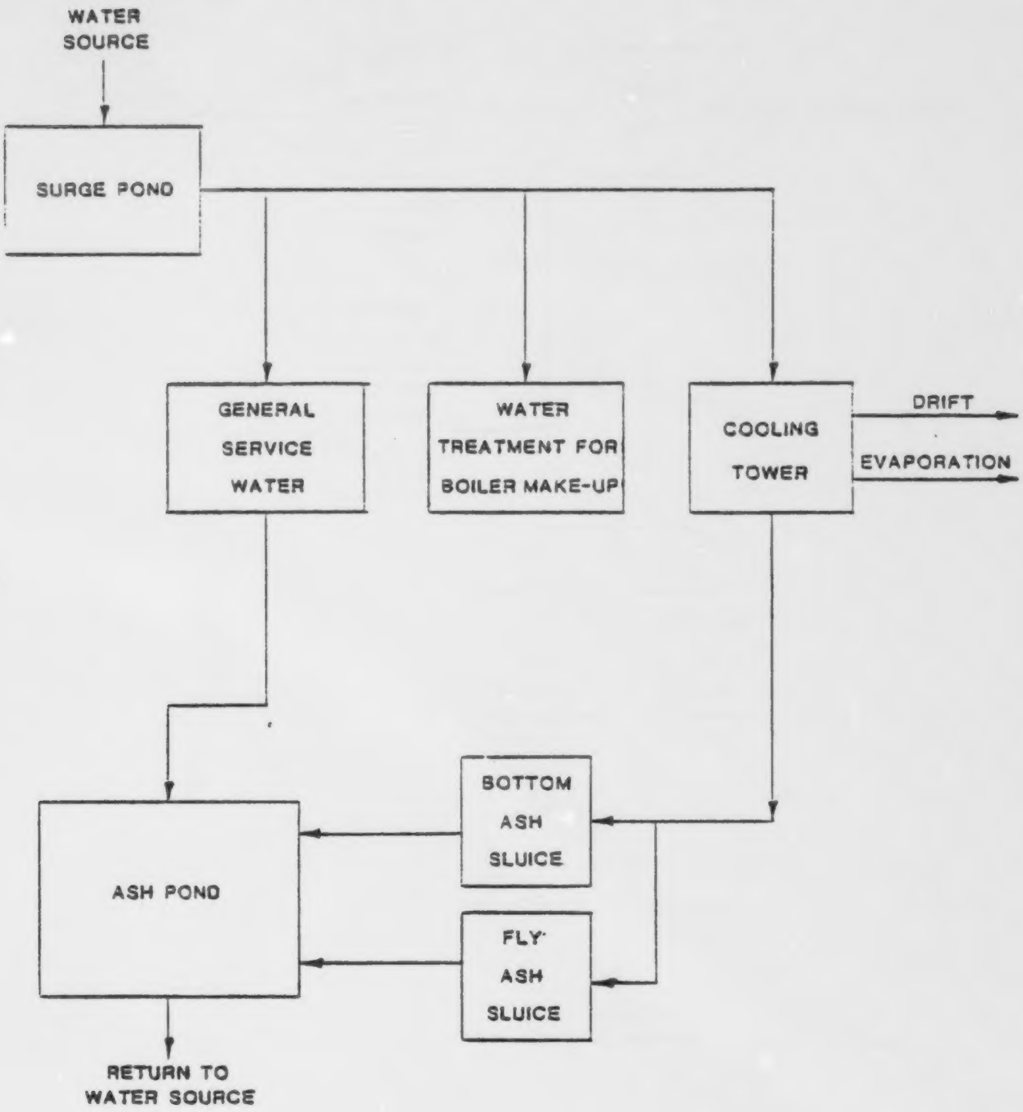
WATER SOURCE
SURGE POND
GENERAL SERVICE WATER
WATER TREATMENT FOR BOILER MAKE-UP
COOLING TOWER
DRIFT
EVAPORATION
BOTTOM ASH SLUICE
FLY ASH SLUICE
ASH POND
RETURN TO WATER SOURCE

Figure 4.1-2. Power plant water system: system #2, partial recirculatory water management.

System #3: Recirculatory cooling at 5.0 cycles of concentration, 50 percent recirculatory ash handling, and recycle of general service water blowdown to the ash handling system; Figure 4.1-3.

System #4: All power plant water systems are recirculatory; Figure 4.1-4.

4.1.1 Cooling Water System

In coal-fired steam/electric power plants, the heat of combustion produces steam to power turbine generators. The steam is subsequently condensed and returned to the boiler for further service. Approximately 45 percent of a fossil-fuel fired generating station's energy is removed and ultimately discharged to the environment by the condenser cooling system (DI-139). To calculate the total cooling water requirement, a power plant efficiency of 37 percent (MC-147) was used. For a 500 MW power plant, 610 MW (35 MM Btu/min) heat removal capacity is required. If a 10°C (20°F) rise in cooling water temperature is assumed in the condenser, a circulating flow of 13 m^3/s (210,000 gpm) is required.

In once-through cooling systems, the makeup water requirement is equal to the circulating rate, i.e. 13 m^3/s (210,000 gpm).

In recirculatory systems, the required blowdown to achieve the desired cycles of concentration is a function of drift (entrained water carried out by the exhaust air) and evaporation:

$$C = \frac{B+D+E}{B+D}$$

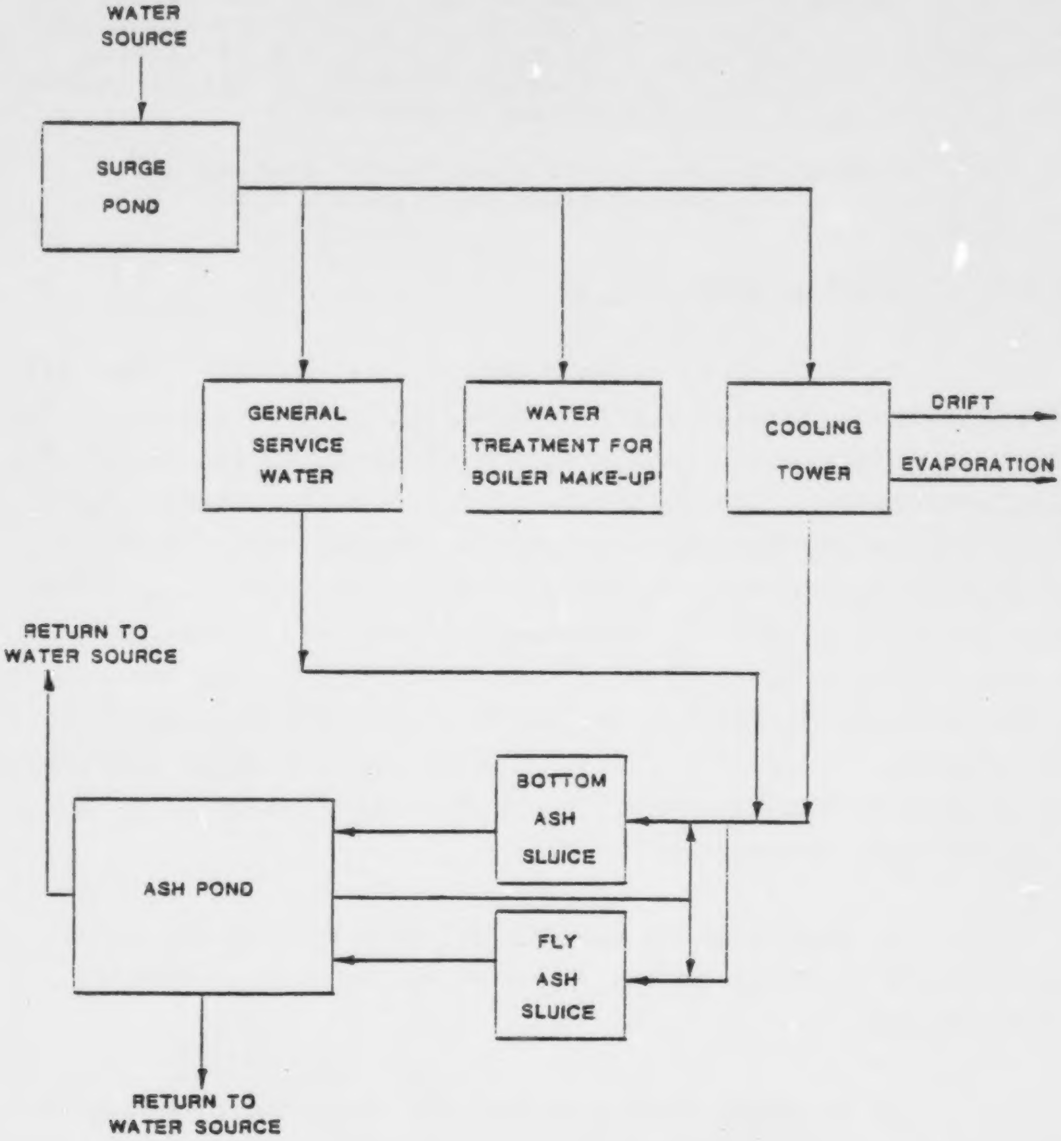
WATER SOURCE
SURGE POND
GENERAL SERVICE WATER
WATER TREATMENT FOR BOILER MAKE-UP
COOLING TOWER
DRIFT
EVAPORATION
RETURN TO WATER SOURCE
BOTTOM ASH SLUICE
ASH POND
FLY ASH SLUICE
RETURN TO WATER SOURCE

Figure 4.1-3. Power plant system: system #3, recirculatory water management.

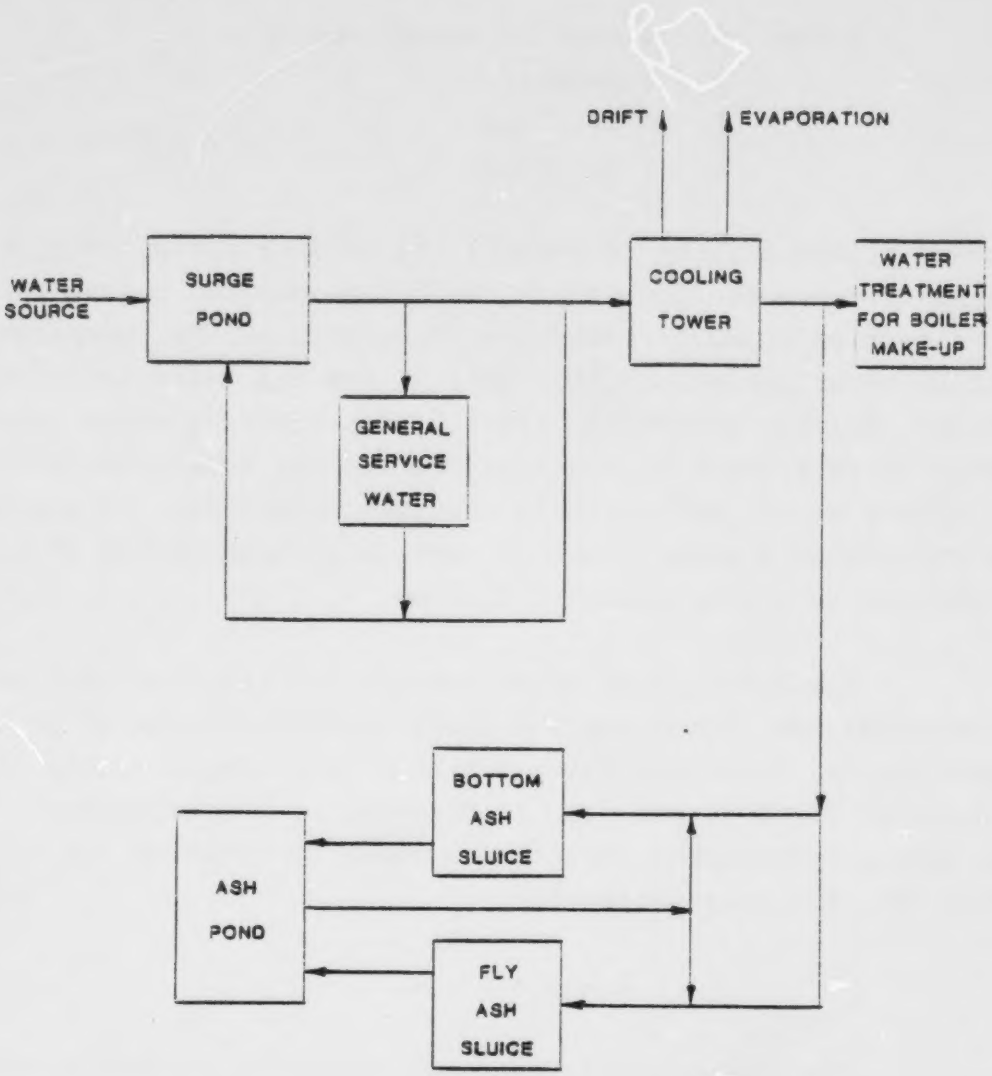
DRIFT
EVAPORATION
WATER SOURCE
SURGE POND
COOLING TOWER
WATER TREATMENT FOR BOILER MAKE-UP
GENERAL SERVICE WATER
BOTTOM ASH SLUICE
ASH POND
FLY ASH SLUICE

Figure 4.1-4. Power plant water system: system #4, zero discharge water management.

where: C = cycles of concentration,
B = blowdown,
D = drift, and
E = evaporation.

A power plants ability to operate its cooling system at high cycles of concentration is limited by the maximum concentration of a limiting impurity (hardness, dissolved solids, suspended solids) or by the solubility limit of scaling salts (calcium sulfate, calcium carbonate, etc.). Recirculating water composition is determined by concentration of the constituents in the makeup water, and possible treatment practices. Therefore, the ability of a power plant to operate at high cycles of concentration is a site-specific problem.

For model plant water consumption calculations, makeup water rates are calculated for three possible ranges of power plant cooling tower operation. Makeup water requirements are calculated for 2.5, 5.0, and 13.5 cycles of concentration. The makeup requirement (M) is determined by blowdown (B), drift (D), and evaporation (E):

$$M = B + D + E.$$

The makeup requirement was extrapolated from plant data from three sites (NO-106, NO-130, CH-387). A general correlation was drawn for makeup requirement vs. cycles of concentration for a 1000 MW power plant (RA-352). This extrapolation assumes that the heat load is proportional to the power plant capacity, and that the evaporation and drift rates are similar. The two data sources showed excellent agreement. It is also assumed that the plant data for evaporation and blowdown rates are characteristic values. The four power plant cooling systems are characterized in Table 4.1-1.

4.1.2 Ash Handling System

In calculating the water requirement for ash handling, Radian has assumed that 1) fly ash is collected in an ESP and wet sluiced to the ash pond as a 5 weight percent slurry, and 2) bottom ash is sluiced to the same pond as a one weight percent slurry. These are characteristic industry values from water recycle/reuse studies that Radian performed for the EPA (NO-106, NO-137, HA-636, GA-203, CH-387). It is also assumed that 75 percent of the ash forms as fly ash and exits in the flue gas, while 25 percent of the ash forms as bottom ash (MC-147). Six representative coals were chosen for the model plant calculations. The compositions are shown in Table A.1-2 (Appendix A). Coal usage rate and ash content determine both the amount of ash to be sluiced and the sluice water requirement. The sluice water requirement for each coal is shown in Table 4.1-2.

Sources for the ash sluicing water vary with the power plant water management systems.

SYSTEM #1: It is assumed that raw water is the only source for the ash sluicing water and that ash sluicing is once-through. Therefore, the water makeup requirement equals the sluicing requirement.

SYSTEM #2: It is assumed that cooling tower blowdown is the source for the ash sluicing water and that ash sluicing is once-through, as shown in Figure 4.1-2. The cooling tower blowdown rate from the previous section is 0.16 m^3/s (2500 gpm).

TABLE 4.1-1. CHARACTERISTIC COOLING SYSTEM OPERATION

System #	Description	Cycles of Concentration	Makeup		Blowdown		Evaporation		Drift	
			m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	Once-through	1	13	(210,000)	13	(210,000)				
2	Partial recirculatory	2.5	0.44	(7,000)	0.16	(2,500)	0.27	(4200)	0.019	(300)
3	Recirculatory	5.0	0.32	(5,000)	0.06	(900)	0.25	(3900)	0.013	(200)
4	Zero Discharge	13.5	0.25	(4,000)	0.02	(300)	0.23	(3670)	0.002	(30)

TABLE 4.1-2. SLUICE WATER REQUIREMENT

Coal	Sluice Water Requirement					
	Fly Ash		Bottom Ash		Total	
	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	0.076	(1200)	0.125	(2000)	0.20	(3100)
2	0.049	(780)	0.082	(1300)	0.13	(2050)
3	0.085	(1350)	0.14	(2250)	0.23	(3600)
4	0.088	(1400)	0.15	(2400)	0.24	(3800)
5	0.045	(720)	0.076	(1200)	0.125	(2000)
6	0.046	(730)	0.076	(1200)	0.13	(2050)

Coals #2, #5, and #6 require less ash sluicing water than available through use of cooling tower blowdown. The excess blowdown could be used for other plant requirements or ponded before discharge. No raw water is required to sluice ash from these coals.

Coals #1, #3, and #4 require more ash sluicing water than available through cooling tower blowdown. It is assumed that this extra requirement is made up with raw water. Cycles of concentration in the cooling tower may also be reduced, or weight percent solids in the slurry could be increased. The makeup water requirement for each coal is shown in Table 4.1-3.

SYSTEM #3: It is assumed that the cooling tower blowdown is one source for ash sluicing water, as shown in Figure 4.1-3. Another assumption requires that the ash sluicing system recycle 50 percent of the sluice water. A second source for sluice water is the collected general services water blowdown (discussed in Section 4.1.3). Table 4.1-4 summarizes the requirements and sources of ash sluice water for each coal.

- With a slight solids increase, coals #2, #5, and #6 can essentially operate without any raw water makeup requirement.

- Coals #1, #3, and #4 require makeup water in the quantities shown in Table 4.1-4. It is probable that a power plant with a water management program of this nature will probably collect "general services" blowdown (discussed in the following section) for use as makeup. Coal #1 can operate without any raw water makeup by a slight solids change. Coals #3 and #4 require raw water makeup in the quantities shown in Table 4.1-4.

TABLE 4.1-3. SYSTEM #2: ASH SLUICE MAKEUP REQUIREMENT

Coal	Total Sluice Requirement m^3/s	Total Sluice Requirement (gpm)	Cooling Tower Blowdown m^3/s	Cooling Tower Blowdown (gpm)	Raw Water Makeup m^3/s	Raw Water Makeup (gpm)
1	0.19	(3100)	0.16	(2500)	0.038	(600)
2	0.13	(2090)	0.13	(2050)	0	0
3	0.23	(3600)	0.16	(2500)	0.063	(1000)
4	0.24	(3800)	0.16	(2500)	0.082	(1300)
5	0.12	(1900)	0.12	(1900)	0	0
6	0.12	(1950)	0.12	(1950)	0	0

TABLE 4.1-4. SYSTEM #3: ASH SLUICE MAKEUP REQUIREMENT

Coal	Total Sluice Water Requirement		Makeup Sluice Water Requirement		Available Cooling Tower Blowdown		Makeup Water Requirement		Available Gen. Serv. Blowdown		Raw Water Requirement	
	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	0.20	(3100)	0.098	(1550)	0.057	(900)	0.041	(650)	0.035	(560)	-	-
2	0.13	(2050)	0.063	(1000)	0.057	(900)	0.0082	(130)				
3	0.23	(3600)	0.11	(1800)	0.057	(900)	0.057	(900)	0.035	(560)	0.021	(330)
4	0.24	(3800)	0.12	(1900)	0.057	(900)	0.063	(1000)	0.035	(560)	0.027	(430)
5	0.12	(1900)	0.060	(950)	0.057	(900)	0.0036	(57)				
6	0.12	(1950)	0.061	(970)	0.057	(900)	0.0046	(73)				

SYSTEM #4: It is assumed that cooling tower blowdown is the source for ash sluicing makeup, as shown in Figure 4.1-4. In this power plant water system, the cooling tower operates at 13.5 cycles of concentration and the ash sluicing system is total recycle. Because the ash settles as a 40-50 percent solids sludge, and assuming a 5 weight percent ash sluice slurry, total recycle translates to 95 percent recycle of the water stream. Table 4.1-5 shows the total sluice water requirement, makeup sluice requirement, and available cooling tower blowdown for each coal.

It can be seen that no additional raw water makeup requirement is necessary for any coal. The excess cooling tower blowdown is available for use as general services water.

4.1.3 General Services Water System

The general services water system is defined to include water conditioning, boiler and condenser cleaning, boiler fireside and air preheater washing, the auxiliary cooling system, and general power plant water use. Data is generally unavailable to define the water consumption for each of these processes and operations. Therefore, to characterize the water requirement for the general services water system, plant data from a water recycle/reuse study (NO-106, CH-387, GA-203) will be used. The data indicates that an assumption of 95 cm^3/s (1.5 gpm) general services water requirement per megawatt will give a reasonably accurate number. Thus for a 500 MW power plant, the general services water requirement is 0.047 m^3/s (750 gpm). Data from the Georgia Power Company, Plant Bowen (NO-106) indicate that approximately 75 percent of the general services water could be available for use as cooling tower makeup or ash sluicing makeup. This rate is 0.035 m^3/s (560 gpm). Table 4.1-6 shows the general services makeup water requirement for each of the systems discussed below:

TABLE 4.1-5. SYSTEM #4: ASH SLUICE MAKEUP REQUIREMENT

Coal	Total Sluice Water Requirement		Makeup Sluice Water Requirement		Available Cooling Tower Blowdown	
	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	0.20	(3100)	0.0095	(150)	0.019	(300)
2	0.13	(2050)	0.0063	(100)	0.019	(300)
3	0.23	(3600)	0.011	(180)	0.019	(300)
4	0.24	(3800)	0.012	(190)	0.019	(300)
5	0.12	(1900)	0.0061	(96)	0.019	(300)
6	0.12	(1950)	0.0061	(97)	0.019	(300)

TABLE 4.1-6. GENERAL SERVICES MAKEUP WATER REQUIREMENT

	Makeup Water Requirement m^3/s	(gpm)	Description
System #1	0.047	750	once-through
System #2	0.047	750	once-through
System #3	0.047	750	recirculated to ash sluicing
System #4	0.012	190	recirculated to cooling tower

SYSTEM #1: All flows are once-through with no attempt being made to reuse any waters.

SYSTEM #2: It is assumed that no attempt is made to reuse this water.

SYSTEM #3: The 0.035 m^3/s (560 gpm) recoverable general services water is required for ash sluicing makeup. The general services requirement is 0.047 m^3/s (750 gpm) and the recycle advantage lowers the raw water requirement for ash sluicing.

SYSTEM #4: The recycle advantage is achieved by combining the recovered general services water with the cooling tower makeup.

4.1.4 Boiler Makeup Water Requirement

Blowdown is required to avoid excessive concentration of impurities in the liquid phase in the boiler. The typical blowdown rate for a drum-type steam boiler is 0.1% (AY-007) of the steam generating rate. For a 500 Mw power plant operating at 37 percent efficiency, approximately 570 cm^3/s (9 gpm) makeup water is required.

4.2 Lime Wet Scrubbing Water Consumption

The lime wet scrubbing process requires fresh makeup water for losses due to: 1) evaporation of water in the absorber, and 2) occlusion in the solid waste. Each model plant system requires a separate calculation due to different flue gas composition and mass rate. For purposes of discussion we will assume one of four equivalent scrubbing trains for a 500 MW power plant burning a 3.5% S coal, with 28 MJ/kg (12,000 Btu/lb) average heating value. Assumptions for each other case are discussed in detail in Appendix B.

4.2.1 Evaporation in the Absorber

The evaporative water loss in the absorber is due to the quenching of hot flue gas by adiabatic saturation. Water evaporated from the scrubbing liquor requires makeup. If there is no sensible heat transfer to the scrubbing liquor, the following heat balance equation applies:

$$\dot{M}C_p\Delta T = \dot{M}_w\lambda$$

where $\dot{M}$ = mass flow rate of the flue gas,

C_p = heat capacity of the flue gas,

ΔT = the change in temperature of the flue gas,

$\dot{M}_w$ = mass rate of water evaporated, and

λ = heat of vaporization of water.

To perform the calculation, a final flue gas outlet temperature is assumed, and $\dot{M}_w$ is calculated. A new gas composition is calculated, and compared to the saturated gas composition at the assumed temperature in psychrometric charts. A new outlet temperature is assumed and the calculation is repeated until the gas compositions are equivalent. In the discussion case, the outlet temperature is 52°C (125°F) and 0.0076 m^3/s (120 gpm) of water are evaporated.

4.2.2 Occlusion in the Solid Waste

The lime wet scrubbing system disposes of the scrubbed sulfur as solid waste. Water is lost both as water of hydration and water occluded in the 60 percent solids sludge waste. In the discussion case, 90 percent removal of the sulfur in the flue gas was assumed, as were 25 percent oxidation and 1.05 lime stoichiometry. (Oxidation can vary considerably, but normally ranges from about 0 to 40% (DT-R-051). Twenty-five percent was chosen as an average value. The assumption of lime stoichiometry of 1.05 is based on Radian experience at the Paddy's Run Station of Lewisville Gas & Electric.) The amount of sulfur to be removed was calculated from known flue gas rate and composition. Assuming that 25 percent of the $CaSO_3$ was oxidized to $CaSO_4$, the water of hydration was calculated. With a lime stoichiometry of 1.05 the excess regenerant was calculated. The final solids concentration in the sludge was assumed to be 60%. This solids level can be achieved through clarification followed by ponding and settling, or by vacuum filtration and/or centrifugation. In the discussion case, 1.5 kg/s (200 lb/min) of waste solids with 1.1 kg/s (150 lb/min) of water of hydration and occluded water were produced. The water of hydration produced was approximately 0.13 kg/s (17 lb/min).

4.3 Limestone Wet Scrubbing Water Consumption

The limestone wet scrubbing process requires fresh makeup water for losses due to: 1) evaporation of water in the absorber, and 2) occlusion in the solid waste. Each model plant system requires a separate calculation due to different flue gas composition and mass rate. For purposes of discussion we will assume one of four equivalent scrubbing trains for a 500 MW power plant burning a 3.5% S coal, with 28 MJ/kg (12,000 Gtu/lb) average heating value. Assumptions for other cases are discussed in detail in Appendix A.

4.3.1 Evaporation in the Absorber

The evaporative water loss in the absorber is due to the quenching of hot flue gas by adiabatic saturation. Water evaporated from the scrubbing liquor requires makeup. If there is no sensible heat transfer to the scrubbing liquor, the heat balance equation presented in Section 4.2.1 applies.

To perform the calculation, a final flue gas outlet temperature is assumed, and the mass flow rate of water is calculated. A new gas composition is calculated, and compared to the saturated gas composition at the assumed temperature in psychrometric charts. When a new outlet temperature is assumed, the calculation is repeated until the gas compositions are equivalent. In the discussion case, the outlet temperature is 53°C (128°F); 0.0076 m^3/s (120 gpm) of water are evaporated per train.

4.3.2 Occlusion in the Solid Waste

The limestone wet scrubbing system disposes of the scrubbed sulfur as solid waste. Water is lost both as water of hydration and water occluded in the 60 percent solids sludge waste. In this case, 90 percent removal of the sulfur in the flue gas was assumed, as were 25 percent oxidation and 1.20 limestone stoichiometry. The amount of sulfur to be removed was calculated from known gas rate and composition. Assuming that 25 percent of the $CaSO_3$ was oxidized to $CaSO_4$, the water of hydration was calculated. The excess regenerant was calculated with a limestone stoichiometry of 1.20. A final solids concentration of 60 percent was assumed in the sludge. This concentration can be achieved through clarification followed by ponding and settling, or by vacuum filtration and/or centrifugation. In the discussion case 1.7 kg/s (220 lb/min) of waste solids with 0.13 kg/s (17 lb/min) water of hydration, and 1.1 kg/s (150 lb/min) occluded water were produced. The total water loss was 1.25 kg/s (165 lb/min) per train.

4.4 Wellman-Lord Sulfite Scrubbing Process Water Consumption

The Wellman-Lord process requires fresh makeup water for losses associated with: 1) evaporation in the prescrubber, 2) sluicing of particulates, 3) drying of the purge solids, 4) water content in the SO_2 product stream, and 5) condenser cooling water blowdown. The largest water loss is due to evaporation in the prescrubber. Particulate slurry and condenser cooling makeup requirements are an order of magnitude less. Drying losses and water in the product stream are again another order of magnitude less. Because calculation of each of these losses is dependent on the model plant case, a separate calculation is required for each. For the purposes of discussion we will assume a base case of one of four equivalent scrubbing trains for a 500 MW power plant burning a 3.5% S coal, with 28 MJ/kg (12,000 Btu/lb) average heating value. Assumptions for each other case are discussed in detail in Appendix A.

4.4.1 Evaporation in the Scrubber

The Wellman-Lord process requires that all particulates and chlorides are removed from the flue gases before entering the absorber. This is commonly accomplished by prescrubbing the gases in a venturi scrubber. The flue gases are cooled from approximately 150°C (300°F) to 50°C (125°F) by adiabatic saturation. The evaporative scrubber loss occurs in the prescrubber rather than the absorber. It is assumed that only water is evaporated by the hot flue gas. There is no sensible heat transfer to the scrubbing liquor. Therefore, the heat balance equation presented in Section 4.2.1 applies.

To perform the calculation, a final flue gas outlet temperature is assumed, and the mass flow rate of water is calculated. A new gas composition is calculated, and compared to the saturated gas composition at the assumed temperature in psychrometric charts. When a new outlet temperature is assumed, the calculation is repeated until the gas compositions are equivalent. In the discussion case the outlet temperature is approximately 53°C (128°F); 0.0076 m^3/s (120 gpm) of water are evaporated per train.

4.4.2 Particulate Sluicing Requirement

A blowdown stream is necessary in the prescrubbing loop to maintain desired circulating suspended and dissolved solids concentrations. A chlorine balance was performed to determine the required blowdown to maintain a dissolved solids concentration of 20,000 mg/l (ppm). The required blowdown to maintain particulates at five weight percent was also calculated. It was found that the desired level of suspended solids (particulates) concentration was the blowdown limiting factor. For the discussion case, 0.83 kg/s (110 lb/min) of water were required for makeup to replace blowdown water in each train.

4.4.3 Water Loss Association with Purge Solids Drying

From water balance calculations, 0.006 kg/s (11.4 lb/min) of purge solids were delivered to the dryer. The centrifuge cake was assumed to contain 0.084 kg/s (11.1 lb/min) of water based on calculations performed by Radian for a comparative FGD systems evaluation (OT-051). When the cake is dried, 100 percent of the water evaporates. When the dryer combustion gases and the water vapor are routed to the prescrubber, the water vapor exits the system with the flue gases. The hot dryer gases evaporate additional water in the prescrubber. The amount of this water loss is calculated on the basis of mass rate of required combustion gases (MC-147) and adiabatic saturation. The total water loss for the discussion case is 0.53 kg/s (70 lb/min) per train.

4.4.4 Water in SO_2 Product Stream

It is assumed that the product SO_2 stream contains 10 weight percent water (OT-051). The amount of SO_2 in the product stream is calculated based on 90 percent removal of the SO_2 in the absorber and 92 percent evolution of SO_2 from the sulfur in the coal (MC-147). The water was calculated as 10 weight percent of the gas stream or 0.072 kg/s (9.5 lb/min) per train.

4.4.5 Condenser Cooling Water Blowdown

In the regeneration loop, the absorber effluent is sent to double effect evaporators. The overhead from the first evaporator passes through a heat exchanger to provide heat for the second effect. If it assumed that a condenser cooling water is required to condense 50 percent of the overhead from both effects, 0.155 m^3/s (2450 gpm) of cooling water are required in the discussion case. This assumes a 10°C (20°F) rise in cooling water temperature in the condenser. A cooling tower operating at five cycles of concentration would require 0.0035 m^3/s (56 gpm) of makeup water, and 0.0007 m^3/s (10.5 gpm) of wastewater would be blowndown per train.

4.5 Magnesia Slurry Absorption Process Water Consumption

The magnesia slurry absorption process requires fresh water makeup for losses due to: 1) evaporation of water in the prescrubber, 2) sluicing of particulates, and 3) water losses associated with drying the magnesium sulfite in the regeneration loop. The largest water loss is due to evaporation in the prescrubber. Both the sluicing requirements and the loss in the dryers are approximately an order of magnitude smaller. Because calculation of each of these losses is dependent on the model plant case, a separate calculation is required for each. For the purpose of discussion we will assume a base case of one of four equivalent scrubbing trains for a 500 MW power plant burning a 3.5 percent S coal, with 28 MJ/kg (12,000 Btu/lb) average heating value. Assumptions for each other case are discussed in detail in Appendix A.

4.5.1 Evaporation in the Scrubber

The magnesia slurry absorption process requires that all particulates and chlorides are removed from the flue gases before entering the absorber. This is commonly accomplished by prescrubbing the gases in a venturi scrubber. The flue gases are cooled from approximately 150°C (300°F) to 50°C (125°F) by adiabatic saturation. The evaporative scrubber loss occurs in the prescrubber rather than the absorber. It is assumed that only water is evaporated by the hot flue gas. There is no sensible heat transfer to the scrubbing liquor. Therefore, the heat balance equation presented in Section 4.2.1 applies.

To perform the calculation, a final flue gas outlet temperature is assumed, and the mass flow rate of water is calculated. A new gas composition is calculated, and compared to the saturated gas composition at the assumed temperature in psychrometric charts. A new outlet temperature is assumed, and the calculation is repeated until the gas compositions are equivalent. In the discussion case the outlet temperature is approximately 53°C (128°F); 0.0075 m^3/s (120 gpm) of water are evaporated per train.

4.5.2 Particulate Sluicing Requirement

A blowdown stream is necessary in the prescrubbing loop to maintain desired circulating suspended and dissolved solids concentrations. A chlorine balance was performed to determine the required blowdown to maintain a dissolved solids concentration of 20,000 mg/l (ppm). The blowdown required to maintain particulates at five weight percent was also calculated. It was found that the desired level of suspended solids (particulates) concentration was the blowdown limiting factor. For the discussion case 0.83 kg/s (110 lb/min) of water were required for makeup to replace blowdown water per train.

4.5.3 Water Losses Associated With Drying

In the magnesia slurry absorption process, a bleedstream containing approximately 10 percent solids is passed through screens for thickening to 40 percent solids. The $MgSO_3 \cdot 6H_2O$ is thermally converted to $MgSO_3 \cdot 3H_2O$ and centrifuged to 95 percent solids. For the discussion case 2.0 kg/sec (260 lb/min) of $MgSO_3 \cdot 3H_2O$ and 0.11 kg/s (15 lb/min) of water are fed to the dryer. In the dryer, the water of hydration and free water evaporate. The water exits with the dryer combustion gases and is sent to the stack. The water loss is 0.71 kg/sec (94 lb/min) per train.

4.6 Double Alkali Wet Scrubbing Water Consumption

The double alkali wet scrubbing process requires fresh makeup water for losses due to: 1) evaporation of water in the absorber, and 2) occlusion in the solid waste. If a prescrubber is required to remove chlorides, an additional water loss would be incurred from the blowdown stream necessary to maintain desired circulating concentrations of suspended and dissolved solids in the prescrubber loop. The water loss due to evaporation in the absorber is the largest water loss. The water loss due to occlusion in the solid waste is approximately an order of magnitude less, as is the prescrubber blowdown if required. Since calculation of each of these losses is dependent on the model plant case, a separate calculation is required for each. For the purpose of discussion we will assume a base case of one of four equivalent scrubbing trains for a 500 MW power plant burning a 3.5% S coal, with 28 MJ/kg (12,000 Btu/lb) average heating value. Assumptions for each other case are discussed in detail in Appendix A.

4.6.1 Evaporation in the Absorber

For application to high chloride coals, the double alkali wet scrubbing process may require that chlorides be removed from the flue gases by a prescrubber before entering the absorber. Under normal conditions, however, a prescrubber will not be used and the evaporative loss occurs in the absorber or in a presaturation chamber in the absorber. It is assumed that only water is evaporated by the hot flue gas. There is no sensible heat transfer to the scrubbing liquor. Therefore, the heat balance equation presented in Section 4.2.1 applies.

To perform the calculation, final flue gas outlet temperature is assumed, and the mass flow rate of water is calculated. A new gas composition is calculated and compared to the saturated gas composition at the assumed temperature in psychrometric charts. A new outlet temperature is assumed, and the calculation is repeated until the gas compositions are equivalent. In the discussion case, the outlet temperature is approximately 53°C (128°F) and 0.0076 m^3/s (120 gpm) of water are evaporated per train.

4.6.2 Occlusion in the Solid Waste

The double alkali wet scrubbing system disposes the scrubbed sulfur as solid waste. Water is lost both as water of hydration and water occluded in the 60 percent solids sludge waste. In the discussion case, 90 percent removal of the sulfur in the flue gas was assumed as were 25 percent oxidation and 1.05 lime stoichiometry. The amount of sulfur to be removed was calculated from known flue gas rate and composition. Assuming that 25 percent of the $CaSO_3$ was oxidized to $CaSO_4$, the water of hydration was calculated.

Excess lime and limestone were calculated for each regenerant. A final solids concentration of 60 percent was assumed in the sludge. In the discussion case, 1.5 kg/s (200 lb/min) of waste solids were produced with lime regenerant, and 1.7 kg/s (220 lb/min) of waste solids were produced with limestone regenerant. The water losses were 1.1 kg/s (150 lb/min) with lime regenerant and 1.25 kg/s (165 lb/min) with limestone regenerant per train.

4.7 Physical Coal Cleaning Water Consumption

It is inherently difficult to describe a generalized coal cleaning process because a large number of processes and operations can be used in common processing areas. It is even more difficult to describe those processes optimized to remove pyritic sulfur. Sulfur in this state is present in many particle sizes and forms, which may be inherent or easily removed. The processes to be used and the extent to which the coal is crushed are dependent upon the amount of sulfur to be removed and its form within the particular coal. Because of the wide variability in physical coal cleaning plant layouts and ranges of operations, it is possible to characterize water consumption only in a general sense. It is possible to attribute water losses to certain specific causes such as drying operations and occlusion in solid wastes. However, these directly attributable losses generally indicate minimum losses. Therefore, plant data presented in Coal Preparation (LE-218) were used to characterize water makeup requirements for coal cleaning plants. These data indicate that in 1962, water consumption in coal preparation plants in the United States averaged 3.3 percent of the circulating water flow rate. New plants are designed to operate closed loop (LE-218). It is impossible to ascertain the extent to which improvements in water consumption may have been made. The water consumption varies from approximately 1.5 percent to 27 percent in the values for various states, with most values being between 1.5 percent and 5.0 percent. Water balance calculations have shown that minimum losses due to occlusion in waste coal and thermal drying or increased coal moisture content are in the range of 1-2 percent of the circulating flow rate (Appendix A). Thus, if it is assumed that modern coal preparation plants operate with water consumption averaging 3.3 percent of the circulating flow, this may be reasonable in regard to realistic average water management practices. With this assumption, the water makeup requirement for the base case 500 MW power plant burning coal that was initially 3.5 percent sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb) is 15 kg/sec (2000 lb/min).

4.8 SO_2 Conversion Processes Water Consumption

Two of the five FGD systems in our study produce a concentrated SO_2 product stream, thus requiring either liquefaction of the SO_2, or conversion into elemental sulfur or sulfuric acid. Both product SO_2 streams can be converted to either elemental sulfur or sulfuric acid. However, for tabulation of water requirements it will be assumed that the magnesium slurry product SO_2 stream will be converted to sulfuric acid, while the Wellman-Lord product stream will be converted to elemental sulfur (as described in Detailed Cost Estimates of Advanced Effluent Desulfurization Processes (MC-147)).

4.8.1 Sulfuric Acid Production

Makeup water is required in sulfuric acid production for: 1) stoichiometric requirement, and 2) cooling water blowdown. The SO_2 stream is oxidized to SO_2 over a vanadium pentoxide catalyst. The SO_2 combines with water in an absorber to form the product acid. The acid is cooled and sent to storage. The stoichiometric requirement is calculated on the basis of one mole of water per mole of SO_2 sent to conversion. Thus 0.75 kg/s (99 lb/min) of water are required for the total 500 MW discussion case conversion SO_2 stream. McGlammery, et al (MC-147) state that the sulfuric acid cooling water requirement is 0.30 m^3/s (4750 gal/min) for the discussion case 500 MW power plant. With a cooling tower operating at five cycles of concentration, 0.0069 m^3/s (110 gal/min) makeup water are required and 0.00013 m^3/s (20.5 gal/min) of wastewater are blown down.

4.8.2 Elemental Sulfur Production

Although several water systems exist in an elemental sulfur production unit, the water requirement is insignificant. Steam is produced in the sulfur condensers at a rate of approximately 1.7 kg/s (225 lb/min) (MC-147) for the compressor seals. The blowdown from these closed loop operations is insignificant in comparison to the other FGD water requirements.

4.9 Model Systems Makeup Water Requirement

The NSPS for SO_2 emissions from coal-fired steam generating plants is currently under review by OAQPS. This review is considering the comprehensive impacts of the existing NSPS and two alternative revised standards. The existing NSPS allows an emission rate of 0.52 $\mu g\ SO_2/J$ (1.2 lb SO_2/MM Btu) of heat input. One alternative standard requires 0.215 $\mu g\ SO_2/J$ (0.5 lb SO_2/MM Btu) of heat input. This standard has the same form as the existing NSPS and thus allows a credit for physical coal cleaning or use of low sulfur coal. The second alternative standard requires 90 percent removal of SO_2 from stack gases, regardless of original sulfur content in the coal.

In order to assess all the various impacts so that a comprehensive conclusion may be drawn, 108 model plant systems have been chosen by the OAQPS as a common base for evaluations. These systems are listed in Table 4.9-1. Generally, these cases will allow analysis of the impacts of the three standards. The type of FGD system, sulfur content of coal, size of steam generator, and degree of coal cleaning will be the variables examined.

> "The first alternative standard deals with the existing NSPS and serves as a baseline for comparing impacts for plants ranging in size from 25 to 1000 megawatts when burning coals with average sulfur contents of 3.5 and 7 percent. It also provides a basis for comparing the impact of a revised standard on eastern and western plants burning typical low sulfur western coal and on a plant using flue gas desulfurization (FGD) in conjunction with coal washing. The second alternative standard represents a 90 percent removal of SO_2 by FGD on all plants regardless of the sulfur content of the coal burned. It does not give any credit for coal cleaning. The third alternative standard has the same format as the existing NSPS; that is, it is based on a fixed mass emission rate. Therefore, plants can use combinations of FGD and coal cleaning, and total removal efficiency will vary depending on the sulfur content of the coal. The third alternative standard is similar in stringency to the second alternative standard in that emissions of 0.4 to 0.5 lb of SO_2/mm Btu represent about 90 percent SO_2 removal efficiency on a typical 3.5 percent sulfur coal.
>
> To limit the scope of work but still provide comparative information, the five viable FGD systems (lime, limestone, magnesium oxide, double alkali, and Wellman-Lord) are considered only in Cases 1(a) and 2(a). This is recommended so that the variations in impacts of the alternative FGD systems will be shown and to provide basic information needed to answer questions which will certainly arise during revision of the NSPS. For example, use of a limestone FGD system results in a sludge disposal problem while use of the magnesium oxide FGD does not result in a sludge. The remainder of the analysis for the recommended alternative is limited to the lime/limestone systems which are the predominant systems used by the domestic industry and which, due to cost, will probably continue to be the first choice of most of the domestic industry for the near future." (CU-077).

4.9.1 Base Uncontrolled Power Plant Water Requirements for the Model Plant Systems

To characterize the range of current water consumption patterns, Radian has chosen to characterize four model power plant water management systems. The four systems are:

TABLE 4.9-1. EPA/OAQPS ALTERNATIVE CONTROL SYSTEMS FOR MODEL PLANTS

Plant sizes to be considered, MW	FGD Systems to be considered	Alternative Standards and Model Plant Systems
25; 100; 500; 1000	5[a]	1. The existing NSPS of 0.52 μg SO_2/J (1.2 lb SO_2/MM Btu) heat input. a. ∿80 percent SO_2 removal on a plant burning a typical coal of 3.5 percent sulfur.
25; 100; 500; 1000	Lime/limestone	b. A plant burning a typical 7 percent sulfur coal with about 90 percent SO_2 removal by FGD.
25; 500	-	c. High and low heating value western and eastern low sulfur coals without FGD for a typical eastern plant.
25; 500	-	d. High and low heating value western low sulfur coals without FGD for a typical western plant.
25; 100; 500; 1000	5[a]	2. a. 90 percent SO_2 removal by FGD on a typical coal of 3.5 percent sulfur and a typical coal of 7 percent sulfur.
25; 100; 500	Lime/limestone	b. 90 percent SO_2 removal by FGD on a plant burning typical high and low heating value western coals of 0.8 percent S (western plant).
25; 500	Lime/limestone	3. 0.215 μg SO_2/J (0.5 lb SO_2 emission/MM Btu) heat input.
		a. 70 to 75 percent SO_2 removal on a plant burning typical high and low heating value western coals of 0.8 percent S (western plant).
25; 500	Lime/limestone	b.1 40 percent sulfur removal by coal washing of a 3.5 percent sulfur coal and 85 percent removal by FGD.
		b.2 40 percent sulfur removal by coal washing of a 7 percent sulfur coal and 95 percent removal by FGD.

[a] The five systems to be considered are lime, limestone, magnesium oxide, double alkali, and Wellman-Lord.

System #1: All power plant water systems are once-through.

System #2: Recirculatory cooling at 2.5 cycles of concentration of once-through ash handling, and once-through general services water.

System #3: Recirculatory cooling at 5.0 cycles of concentration, 50 percent recirculatory ash handling, and recycle of general service water blowdown to the ash handling system.

System #4: All power plant water systems are recirculatory.

Each of the process water requirements and methods of their calculation have been previously discussed in this section. Table 4.9-2 summarizes the results of the calculations for Systems #1 through #4 for the discussion case 500 MW power plant. In the once-through system, since the cooling water requirement dominates, FGD water requirements are insignificant by comparison. In recirculatory systems operating at 2.5, 5.0, and 13.5 cycles of concentration, ash handling, general services, and FGD water requirements become significant. System #3 has been chosen as representative of "typical" power plant water requirements for comparison to FGD requirements. This system may have lower water requirements than power plant systems in current operation, but it will become more predominant as the national zero discharge goal nears. This system characterizes water requirements in the midrange of the three recirculatory systems. Model systems base power plant water requirements calculated for System #3 are shown in Table 4.9-4.

TABLE 4.9-2. BASE CASE: MODEL POWER PLANT WATER CONSUMPTION

System Number	System[a] Description	Cooling Water System		Ash Handling System		General Services Water		Boiler Makeup		Total Fresh Water Makeup Requirement	
		m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	Once-through[b]	13	(210,000)	0.23	(3600)	0.047	(750)	0.0006	(9)	13.5	(215,000)
2	Partial Recirculatory[c]	0.44	(7,000)	0.069	(1100)	0.047	(750)	0.0006	(9)	0.56	(8,850)
3	Recirculatory[d]	0.32	(5,000)	0.021	(330)	0.047	(750)	0.0006	(9)	0.38	(6,100)
4	Zero Discharge[e]	0.25	(4,000)	0	(0)	0.012	(190)	0.0006	(9)	0.26	(4,200)

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[a] The base case is a 500 MW power plant operating at an efficiency of 37%; 3.5% S coal; average heating value of 28 MJ/kg (12,000 Btu/lb).

[b] All power plant water systems are once-through; refer to Figure 4.1-1.

[c] Recirculatory cooling at 2.5 cycles of concentration, once-through ash handling, and once-through general services water; refer to Figure 4.1-2.

[d] Recirculatory cooling at 5.0 cycles of concentration, 50% recirculatory ash handling, and recycle of general service water blowdown to the ash handling system; refer to Figure 4.1-3.

[e] All power plant water systems are recirculatory; refer to Figure 4.1-4.

4.9.2 SO_x Control Strategy Water Requirements for the Model Plant Systems

Each of the SO_x control strategy water requirements and methods of calculation have been discussed in previous sections. Table 4.9-3 summarizes the results of each of these calculations for the base case 500 MW power plant. This table illustrates that the method used to calculate the total FGD process water requirement was to define and calculate specific requirements with their sum being the total requirement. The most significant requirement is evaporative loss in the scrubber. Cooling water requirements for the Wellman-Lord Sulfite Scrubbing Process and SO_2 conversion to sulfuric acid are also significant. Other significant requirements are approximately an order of magnitude less.

4.9.3 Matrix Presentation of Model Plant Water Requirements

Table 4.9-4 presents the results of water requirement calculations for each model plant. A detailed discussion of the methods and assumptions used in calculating these requirements are given in Appendix A.

TABLE 4.9-3. BASE CASE: FGD SYSTEM WATER CONSUMPTION[h]

Water[a] Requirement	Lime Wet Scrubbing		Limestone Wet Scrubbing		Wellman-Lord Sulfite Scrubbing		Magnesia Slurry Absorption		Double Alkali Wet Scrubbing			
									Lime Regenerant		Limestone Regenerant	
	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
Evaporative loss,	0.030	(480)	0.030	(480)	0.030	(480)	0.030	(480)	0.030	(480)	0.030	(480)
Loss in solid waste,	0.005	(80)	0.005	(80)					0.005	(72)	0.005	(79)
Prescrubber blowdown,					0.003	(54)	0.003	(54)				
Cooling water blowdown,					0.014	(220)						
Loss with solids drying,					0.002	(34)	0.003	(45)				
Loss in product SO_2 stream,					0.003	(5)						
SO_2 Conversion Requirement:												
Sulfuric acid					0.008	(130)	0.008	(130)				
Elemental Sulfur					0.002	(3)	0.0002	(3)				
Total	0.035	(560)	0.035	(560)	0.058	(923)[d]	0.024	(710)[b,d]	0.035	(552)[c]	0.035	(559)[c]
					0.050	(795)[b,e]	0.036	(580)[e]				
Liquid to gas ratio required for scrubber - m^3/Nm^3 (gal/1000 scf)	.005-.015	(35-110)	.005-.015	(35-110)	0.004	(3)[f]	.003-.004	(20-30)	.0007-.002	(5-15)	0.007-.002	(5-15)
					.002	(15)[g]	.002	(15)[g]				

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[a] The base case is for a 500 MW power plant operating at an efficiency of 37%, burning 3.5% burning 3.5% sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb).

[b] This value is assumed for model plant calculations.

[c] The average of the two double alkali systems is used for model plant calculations.

[d] This is the total water requirement if sulfuric acid is produced.

[e] This is the total water requirement if sulfur is produced.

[f] Separate scrubbing loops are provided for each of 3 trays.

[g] Prescrubber.

[h] Make-up water requirement.

TABLE 4.9-4. MODEL PLANT SYSTEM WATER REQUIREMENTS

Case	Power Plant Capacity MW	SO$_x$ Control Strategy	% Sulfur Removal	Coal Type[a]	System #3 Power Plant Makeup Water Requirement		SO$_x$ Control Strategy Makeup Water Requirement		Total Model System Makeup Water Requirement	
					m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
1	1000	Lime	76	#3	0.76	(12,000)	0.069	(1,100)	0.82	(13,000)
2	1000	Limestone	76	#3	0.76	(12,000)	0.069	(1,100)	0.82	(13,000)
3	1000	Wellman-Lord	76	#3	0.76	(12,000)	0.095	(1,500)	0.88	(14,000)
4	1000	Magnesium Oxide	76	#3	0.76	(12,000)	0.0[illegible]	(1,300)	0.82	(13.000)
5	1000	Double Alkali	76	#3	0.76	(12,000)	0.069	(1,100)	0.82	(13,000)
6	500	Lime	76	#3	0.38	(6,100)	0.034	(540)	0.42	(6,600)
7	500	Limestone	76	#3	0.38	(6,100)	0.034	(540)	0.42	(6,600)
8	500	Wellman-Lord	76	#3	0.38	(6,100)	0.048	(760)	0.43	(6,800)
9	500	Magnesium Oxide	76	#3	0.38	(6,100)	0.042	(670)	0.43	(6,800)
10	500	Double Alkali	76	#3	0.38	(6,100)	0.034	(540)	0.42	(6,600)
11	100	Lime	76	#3	0.076	(1,200)	0.0069	(110)	0.082	(1,300)
12	100	Limestone	76	#3	0.076	(1,200)	0.0069	(110)	0.082	(1,300)
13	100	Wellman-Lord	76	#3	0.076	(1,200)	0.0095	(150)	0.088	(1,400)
14	100	Magnesium Oxide	76	#3	0.076	(1,200)	0.0088	(140)	0.088	(1,400)
15	100	Double Alkali	76	#3	0.076	(1,200)	0.0069	(110)	0.082	(1,300)
16	25	Lime	76	#3	0.019	(300)	0.0017	(27)	0.021	(330)
17	25	Limestone	76	#3	0.019	(300)	0.0017	(27)	0.021	(330)
18	25	Wellman-Lord	76	#3	0.019	(300)	0.0024	(38)	0.021	(340)
19	25	Magnesium Oxide	76	#3	0.019	(300)	0.0023	(37)	0.021	(340)
20	25	Double Alkali	76	#3	0.019	(300)	0.0017	(27)	0.021	(330)
21	1000	Lime	88	#4	0.76	(12,000)	0.079	(1,250)	0.82	(13,000)
22	1000	Limestone	88	#4	0.76	(12,000)	0.082	(1,300)	0.82	(13,000)
23	500	Lime	88	#4	0.39	(6,200)	0.039	(620)	0.43	(6,800)
24	500	Limestone	88	#4	0.39	(6,200)	0.040	(640)	0.43	(6,800)
25	100	Lime	88	#4	0.76	(1,200)	0.0082	(130)	0.082	(1,300)
26	100	Limestone	88	#4	0.76	(1,200)	0.0082	(130)	0.082	(1,300)
27	25	Lime	88	#4	0.020	(310)	0.0020	(31)	0.021	(340)
28	25	Limestone	88	#4	0.020	(310)	0.0020	(32)	0.021	(340)

(Continued)

TABLE 4.9-4. MODEL PLANT SYSTEM WATER REQUIREMENTS (Continued)

Case	Power Plant Capacity MW	SO_x Control Strategy	% Sulfur Removal	Coal Type [a]	System #3 Power Plant Makeup Water Requirement		SO_x Control Strategy Makeup Water Requirement		Total Model System Makeup Water Requirement	
					m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
29	500	–	N/A	#1	0.37	(5,800)		N/A	0.37	(5,800)
30	500	–	N/A	#2	0.37	(5,800)		N/A	0.37	(5,800)
31	500	–	N/A	#3	0.38	(6,100)		N/A	0.38	(6,100)
32	25	–	N/A	#1	0.018	(290)		N/A	0.018	(290)
33	25	–	N/A	#2	0.018	(290)		N/A	0.018	(290)
34	25	–	N/A	#3	0.020	(310)		N/A	0.020	(310)
35	500	–	N/A	#1	0.37	(5,800)		N/A	0.37	(5,800)
36	500	–	N/A	#2	0.37	(5,800)		N/A	0.37	(5,800)
37	25	–	N/A	#1	0.018	(290)		N/A	0.018	(290)
38	25	–	N/A	#2	0.018	(290)		N/A	0.018	(290)
39	500	Coal Cleaning/Lime	40/39	#5	0.37	(5,800)	0.049	(770)	0.41	(6,500)
40	500	Coal Cleaning/Limestone	40/39	#5	0.37	(5,800)	0.049	(770)	0.41	(6,500)
41	1000	Lime	90	#3	0.76	(12,000)	0.070	(1,100)	0.82	(13,000)
42	1000	Limestone	90	#3	0.76	(12,000)	0.070	(1,100)	0.82	(13,000)
43	1000	Wellman-Lord	90	#3	0.76	(12,000)	0.095	(1,500)	0.88	(14,000)
44	1000	Magnesium Oxide	90	#3	0.76	(12,000)	0.088	(1,400)	0.82	(13,000)
45	1000	Double Alkali	90	#3	0.76	(12,000)	0.070	(1,100)	0.82	(13,000)
46	500	Lime	90	#3	0.38	(6,100)	0.035	(550)	0.42	(6,600)
47	500	Limestone	90	#3	0.38	(6,100)	0.035	(550)	0.42	(6,600)
48	500	Wellman-Lord	90	#3	0.38	(6,100)	0.049	(770)	0.44	(6,900)
49	500	Magnesium Oxide	90	#3	0.38	(6,100)	0.044	(700)	0.43	(6,800)
50	500	Double Alkali	90	#3	0.38	(6,100)	0.035	(550)	0.42	(6,600)
51	100	Lime	90	#3	0.076	(1,200)	0.0069	(110)	0.082	(1,300)
52	100	Limestone	90	#3	0.076	(1,200)	0.0069	(110)	0.082	(1,300)
53	100	Wellman-Lord	90	#3	0.076	(1,200)	0.0095	(150)	0.088	(1,400)
54	100	Magnesium Oxide	90	#3	0.076	(1,200)	0.0088	(140)	0.082	(1,300)
55	100	Double Alkali	90	#3	0.076	(1,200)	0.0069	(110)	0.082	(1,300)
56	25	Lime	90	#3	0.020	(310)	0.0017	(27)	0.021	(340)

(Continued)

TABLE 4.9-4. MODEL PLANT SYSTEM WATER REQUIREMENTS (Continued)

Case	Power Plant Capacity MW	SO_x Control Strategy	% Sulfur Removal	Coal Type[a]	System #3 Power Plant Makeup Water Requirement		SO_x Control Strategy Makeup Water Requirement		Total Model System Makeup Water Requirement	
					m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
57	25	Limestone	90	#3	0.020	(310)	0.0018	(28)	0.021	(340)
58	25	Wellman-Lord	90	#3	0.020	(310)	0.0024	(38)	0.022	(350)
59	25	Magnesium Oxide	90	#3	0.020	(310)	0.0022	(35)	0.021	(340)
60	25	Double Alkali	90	#3	0.020	(310)	0.0017	(27)	0.021	(340)
61	1000	Lime	90	#4	0.76	(12,000)	0.082	(1,300)	0.82	(13,000)
62	1000	Limestone	90	#4	0.76	(12,000)	0.082	(1,300)	0.82	(13,000)
63	1000	Wellman-Lord	90	#4	0.76	(12,000)	0.11	(1,800)	0.88	(14,000)
64	1000	Magnesium Oxide	90	#4	0.76	(12,000)	0.11	(1,700)	0.88	(14,000)
65	1000	Double Alkali	90	#4	0.76	(12,000)	0.82	(1,300)	0.82	(13,000)
66	500	Lime	90	#4	0.39	(6,200)	0.040	(630)	0.43	(6,800)
67	500	Limestone	90	#4	0.39	(6,200)	0.040	(640)	0.43	(6,800)
68	500	Wellman-Lord	90	#4	0.39	(6,200)	0.056	(890)	0.45	(7,100)
69	500	Magnesium Oxide	90	#4	0.39	(6,200)	0.054	(850)	0.44	(7,000)
70	500	Double Alkali	90	#4	0.39	(6,200)	0.040	(630)	0.43	(6,800)
71	100	Lime	90	#4	0.076	(1,200)	0.0082	(130)	0.082	(1,300)
72	100	Limestone	90	#4	0.076	(1,200)	0.0082	(130)	0.082	(1,300)
73	100	Wellman-Lord	90	#4	0.076	(1,200)	0.011	(180)	0.088	(1,400)
74	100	Magnesium Oxide	90	#4	0.076	(1,200)	0.011	(170)	0.088	(1,400)
75	100	Double Alkali	90	#4	0.076	(1,200)	0.0082	(130)	0.082	(1,300)
76	25	Lime	90	#4	0.020	(310)	0.0022	(35)	0.022	(350)
77	25	Limestone	90	#4	0.020	(310)	0.0023	(36)	0.022	(350)
78	25	Wellman-Lord	90	#4	0.020	(310)	0.0031	(49)	0.023	(360)
79	25	Magnesium Oxide	90	#4	0.020	(310)	0.0029	(46)	0.023	(360)
80	25	Double Alkali	90	#4	0.020	(310)	0.0022	(35)	0.022	(350)
81	500	Lime	90	#1	0.37	(5,800)	0.035	(550)	0.40	(6,300)
82	500	Lime	90	#2	0.37	(5,800)	0.034	(540)	0.40	(6,300)
83	500	Limestone	90	#1	0.37	(5,800)	0.035	(550)	0.40	(6,300)
84	500	Limestone	90	#2	0.37	(5,800)	0.034	(540)	0.40	(6,300)

(Continued)

TABLE 4.9-4. MODEL PLANT SYSTEM WATER REQUIREMENTS (Continued)

Case	Power Plant Capacity MW	SO_x Control Strategy	% Sulfur Removal	Coal Type[a]	System #1 Power Plant Makeup Water Requirement		SO_x Control Strategy Makeup Water Requirement		Total Model System Makeup Water Requirement	
					m^3/s	(gpm)	m^3/s	(gpm)	m^3/s	(gpm)
85	100	Lime	90	#1	0.073	(1,150)	0.0069	(110)	0.082	(1,300)
86	100	Lime	90	#2	0.073	(1,150)	0.0069	(110)	0.082	(1,300)
87	100	Limestone	90	#1	0.073	(1,150)	0.0069	(110)	0.082	(1,300)
88	100	Limestone	90	#2	0.073	(1,150)	0.0069	(110)	0.082	(1,300)
89	25	Lime	90	#1	0.018	(290)	0.0017	(27)	0.020	(320)
90	25	Lime	90	#2	0.018	(290)	0.0017	(27)	0.020	(320)
91	25	Limestone	90	#1	0.018	(290)	0.0018	(28)	0.020	(320)
92	25	Limestone	90	#2	0.018	(290)	0.0017	(27)	0.020	(320)
93	500	Lime	70	#1	0.37	(5,800)	0.034	(540)	0.40	(6,300)
94	500	Lime	70	#2	0.37	(5,800)	0.031	(490)	0.40	(6,300)
95	500	Limestone	70	#1	0.37	(5,800)	0.034	(540)	0.40	(6,300)
96	500	Limestone	70	#2	0.37	(5,800)	0.032	(500)	0.40	(6,300)
97	25	Lime	70	#1	0.018	(290)	0.0017	(27)	0.020	(320)
98	25	Lime	70	#2	0.018	(290)	0.0015	(24)	0.020	(310)
99	25	Limestone	70	#1	0.018	(290)	0.0017	(27)	0.020	(320)
100	25	Limestone	70	#2	0.018	(290)	0.0016	(25)	0.020	(320)
101	500	Coal Cleaning/Lime	40/85	#5	0.37	(5,800)	0.050	(790)	0.42	(6,600)
102	500	Coal Cleaning/Limestone	40/85	#5	0.37	(5,800)	0.050	(790)	0.42	(6,600)
103	25	Coal Cleaning/Lime	40/85	#5	0.018	(290)	0.0025	(39)	0.021	(330)
104	25	Coal Cleaning/Limestone	40/85	#5	0.018	(290)	0.0025	(40)	0.021	(330)
105	500	Coal Cleaning/Lime	40/91	#6	0.37	(5,800)	0.052	(820)	0.42	(6,600)
106	500	Coal Cleaning/Limestone	40/91	#6	0.37	(5,800)	0.052	(830)	0.42	(6,600)
107	25	Coal Cleaning/Lime	40/91	#6	0.018	(290)	0.0026	(41)	0.021	(330)
108	25	Coal Cleaning/Limestone	40/91	#6	0.018	(290)	0.0026	(41)	0.021	(330)

[a] coal #1 : 0.8% S; 19 MJ/kg (8,000 Btu/lb); 6% ash; 30% H_2O
coal #2 : 0.8% S; 26 MJ/kg (11,000 Btu/lb); 6% ash; 15% H_2O
coal #3 : 3.5% S; 28 MJ/kg (12,000 Btu/lb); 12% ash; 2.6% H_2O
coal #4 : 7.0% S; 28 MJ/kg (12,000 Btu/lb); 12% ash; 5.7% H_2O
coal #5 : 2.0% S; 26 MJ/kg (11,000 Btu/lb); 6% ash; 15% H_2O
coal #6 : 4.0% S; 27 MJ/kg (11,500 Btu/lb); 6% ash; 15% H_2O

5.0 CHARACTERIZATION OF PROCESS WASTEWATERS

This section assesses the impact of SO_x control technologies on power plant wastewater streams. Wastewaters of power plants uncontrolled for SO_x emissions are characterized. In addition, wastewaters from the five subject FGD systems, the physical coal cleaning process, and two SO_2 conversion processes are characterized. Because water management at steam/electric utilities is highly site-specific, it is necessary to characterize the wastewater streams in a general manner. A quantitative characterization of typical power plant wastewater streams is not possible. Wherever possible, characteristic compositions and quantities of effluent streams are given to indicate typical ranges of operation. Impact is assessed on the basis of a general comparison of effluents. Fortunately, this comparison is easily made, due to the significant impact of uncontrolled power plant wastewaters on the surrounding water quality. The minor nature of the wastewater streams associated with SO_x control technologies is also a contributing factor.

Wastewater streams are characterized for the following processes:

- Coal-fired Power Plant, Uncontrolled for SO_x Emissions
- Lime Wet Scrubbing
- Limestone Wet Scrubbing
- Wellman-Lord Process
- Magnesia Slurry Absorption Process
- Double Alkali Wet Scrubbing
- Physical Coal Cleaning

- Allied Chemical SOx Reduction Process
- Single Absorption Contact Sulfuric Acid Plant

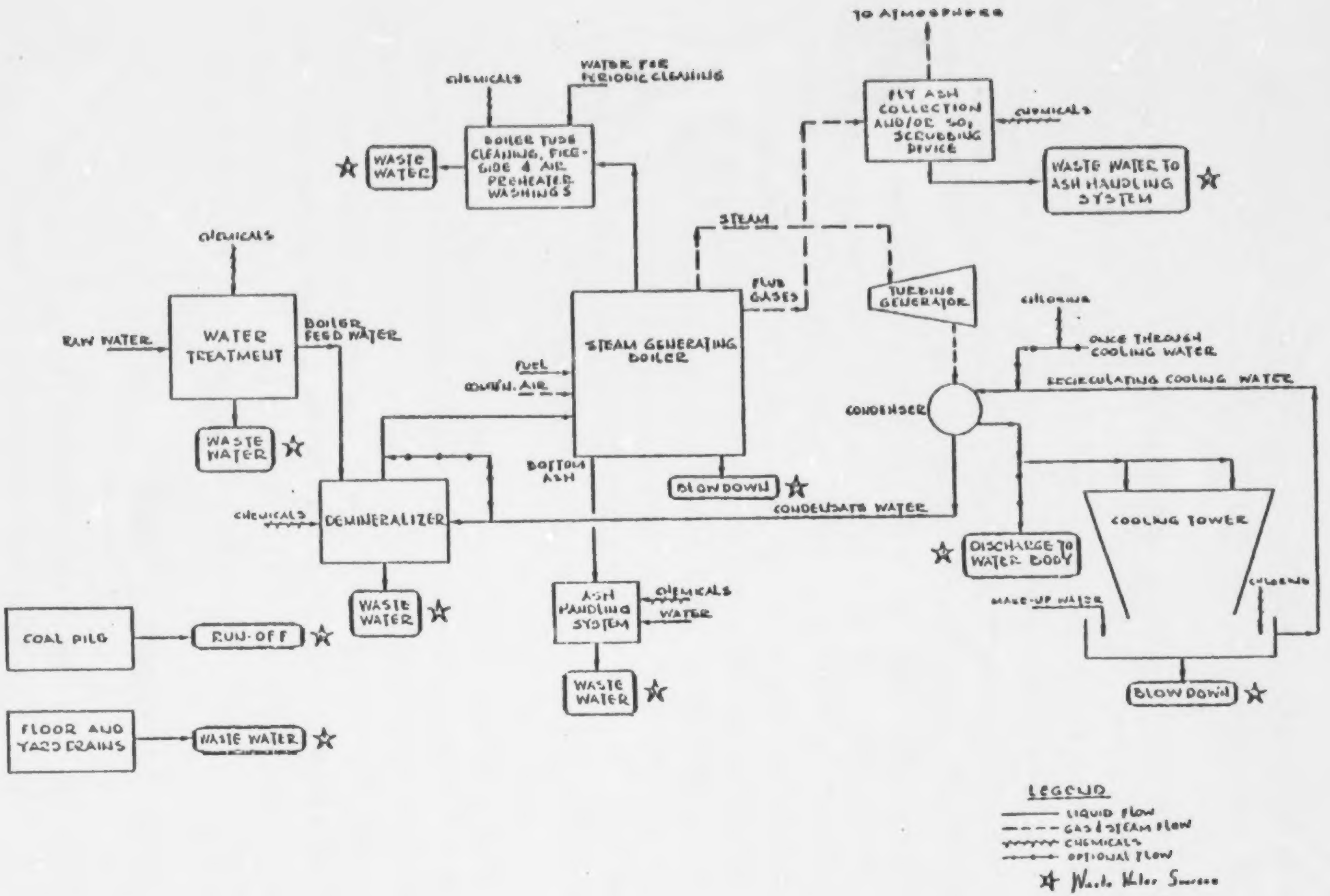
5.1 Characterization of Wastewaters from a Power Plant Not Equipped with an FGD System

Wastewater effluents are discharged from many sources in steam/electric power plants. Because of the wide variability in individual plant layouts, water intake quality, selection and operation of process units, and water management strategies, however, not all plants have the same effluents. Discharges from similar sources at different plants may also have highly dissimilar characteristics in terms of stream flow and composition. This is in part attributable to wide variations in raw feed water quality, ash composition, coal composition, quantity of ash, and selected slate of boiler water and cooling tower additives. There are also several options for handling wastes such as recycle, combining with other waste streams, slip stream or full stream treatment, and discharge of once-through streams. The last option is becoming less practicable, however, due to promulgation of zero discharge standards.

The main point that must be considered in addressing water treatment technologies is that water management at steam/electric utilities is highly site-specific. Attempts to generalize about aqueous discharges and their treatment should be carefully qualified since a single description of any given effluent is not generally applicable to all power plants. In this section a description of all major utility effluents will be presented.

5.1.1 Power Plant Wastewater Sources

Figure 5.1-1 graphically summarizes the sources of wastewater in a fossil fueled generating station and depicts the interrelation of the various processes producing wastewater.


TO ATMOSPHERE
CHEMICALS
WATER FOR PERIODIC CLEANING
WASTE WATER
BOILER TUBE CLEANING, FIRE-SIDE & AIR PREHEATER WASHINGS
FLY ASH COLLECTION AND/OR SO2 SCRUBBING DEVICE
CHEMICALS
WASTE WATER TO ASH HANDLING SYSTEM
STEAM
FLUE GASES
TURBINE GENERATOR
CHLORINE
ONCE THROUGH COOLING WATER
RECIRCULATING COOLING WATER
CHEMICALS
RAW WATER
WATER TREATMENT
BOILER FEED WATER
WASTE WATER
FUEL
COMB'N. AIR
STEAM GENERATING BOILER
CONDENSER
BOTTOM ASH
BLOWDOWN
CONDENSATE WATER
CHEMICALS
DEMINERALIZER
COOLING TOWER
DISCHARGE TO WATER BODY
MAKE-UP WATER
CHLORINE
WASTE WATER
ASH HANDLING SYSTEM
CHEMICALS
WATER
COAL PILE
RUN-OFF
WASTE WATER
BLOWDOWN
FLOOR AND YARD DRAINS
WASTE WATER
LEGEND
LIQUID FLOW
GAS & STEAM FLOW
CHEMICALS
OPTIONAL FLOW
Waste Water Sources

Reference: BU-087

Figure 5.1-1. Sources of wastewater in a fossil-fueled steam-electric plant.

The basic sources of wastewater are:

- cooling water system,
- ash handling,
- water conditioning,
- boiler blowdown,
- coal pile runoff,
- equipment cleaning,
- general plant drainage,
- process spills and leaks, and
- miscellaneous sources.

Flue gas desulfurization wastes will be discussed in the following sections.

The frequency of waste discharge can be any of the following:

> Continuous - Wastewater is discharged at a fairly constant rate without interruption as long as the plant is operating.
>
> Intermittent - Wastewater is discharged on a regular or scheduled basis (*e.g.*, every shift, daily) when the plant is operational.
>
> Periodic - Wastewater is discharged at infrequent intervals (monthly or yearly) which may or may not be of regular frequency.

Significant water parameters which can be used to characterize wastewater streams from fossil-fueled power plants include:

- Suspended Solids - undissolved solid matter which is present in water,
- Dissolved Solids - dissolved matter which is present aqueous solution,
- pH - a measure of the acidity of alkalinity of a waste stream,
- Hardness - the soluble calcium and magnesium content of a waste stream,
- Toxic Chemical Species - aqueous constituents such as heavy metals, certain organic compounds, etc., that are generally present in trace concentrations but which are highly detrimental to receiving bodies of water,
- Oily Wastes - the insoluble organics in water such as oils, greases, etc.
- Biochemical Oxygen Demand - the oxygen required to stabilize soluble and/or insoluble impurities in water by biochemical reaction, and
- Chemical Oxygen Demand - the oxygen required to convert water impurities to their oxidized forms through the action of oxidizing agents.

Identification of trace toxic chemicals in utility effluents is currently underway. Preliminary results of a literature review, utility survey, and preliminary screening sampling study indicate that approximately 31 of the 126 "unambiguous" toxic substances may be found in utility streams. An in-depth sampling program is now in progress under EPA sponsorship to further define these streams.

The possible sources of these pollutants may be any of the following:

Cooling System Treatment
Corrosion and Scale Inhibitor
Corrosion Inhibitor
Biocide - Cooling System
Algacide
Insulation
Maintenance Material
Transformer Fluid
Condensate Neutralizer
Cleaning Product
Ash Constituent
Construction Material
Cooling Tower Material
Lab Reagent
Instrument Use

In the following discussion, each of the wastewater sources potentially present in utilities is described. To the greatest extent possible the composition and flow rate of the streams will be characterized.

Cooling Water System

Approximately forty-five percent of a fossil-fuel fired generating station's energy is removed and ultimately discharged to the environment by the condenser cooling system. Basically, two condenser cooling systems are employed by the electric utility industry: 1) once-through system and 2) recirculating system.

Once-Through Cooling Water

Once-through cooling systems are unique since the total cooling water flow for heat removal is discharged as a wastewater effluent. This cooling water flow rate is approximately 100 cm^3/Kcal of heat removal for every 10°C of cooling water temperature rise (12 gal/1000 Btu of heat removal for every 10°F of cooling water temperature rise). After passing through the condenser, the cooling water is discharged to a heat sink (*i.e.*, river, lake, pond) where the heat is dissipated.

Due to the nature of the once-through system, the chemical composition of the effluent water is essentially equivalent to that of the influent water. Water quality parameters such as total dissolved and suspended solids, pH, etc., are largely governed by the characteristics of the cooling water source, and they are not significantly altered by the operation of the cooling system. Slight changes in the chemical composition between influent and effluent for these systems may occur, however, due to 1) formation of corrosion products and/or 2) addition of treatment chemicals such as chlorine.

Water-side corrosion of the main condenser will result in corrosion products (*i.e.*, metal oxides) appearing in the cooling water effluent. Condenser metallurgy is carefully selected, however, to minimize water-side corrosion rates to the extent that negligible quantities of corrosion products appear in the effluent cooling water (EN-127).

Extensive cooling water treatment is normally precluded in once-through systems due to the large quantities of raw water used. However, chemical treatment with biocides is often necessary to control biological growths of algae and slime that accumulate on condenser surfaces, retard heat transfer, and obstruct cooling water flow. Chlorine is by far the most common biocide used; however, on rare occasions sodium hypochlorite is used instead.

Cooling water is chlorinated by "shock" or "slug" treatment methods in which a large dose of chlorine is added intermittently. The duration of "shock" treatment ranges from five minutes to two hours, but typically lasts 30 minutes. The frequency of chlorine treatment ranges from one to ten times per day (typically once per shift or three times per day) (MA-230). During chlorination, free residual chlorine is kept between 0.1 and 0.2 mg/ℓ (ppm) in the condenser effluent. When using seawater as a coolant, chlorine residuals as high as 12 mg/ℓ (ppm) may be used to deter the presence of eels and jelly fish as well as to inhibit the mussel and crustacean growths on the condenser (MA-230).

Recirculating System: Cooling Tower Blowdown

Recirculating cooling systems employ cooling devices such as cooling towers, spray ponds/canals, etc., which allow the reuse of cooling water. These devices promote cooling primarily by evaporating a portion of the recirculating water flow. Impurities and contaminants that come into the system with makeup water and other sources become concentrated. A blowdown stream is withdrawn from the system to control the concentration of impurities and contaminants. This stream represents the recirculating cooling system wastewater. Blowdown quantity is set by the maximum concentration of a limiting impurity (*i.e.*, hardness, dissolved solids, suspended solids) that can be tolerated in the system or by the solubility limit of scaling salts such as calcium sulfate, calcium carbonate, etc.

(Example limits are 5 weight percent suspended solids and 20,000 mg/ℓ (ppm) dissolved solids.) The blowdown rate typically ranges between 0.5 and 3.0% of the recirculating water flow (DO-051). The recirculating flow is approximately 100 cm^3/Kcal of heat removal for every 10°C of cooling water temperature rise (12 gal/1000 Btu for 10°F).

The blowdown from recirculating cooling systems has the same chemical composition as does the recirculating cooling water. The major factors that influence cooling water composition include:

- Makeup water characteristics
- Chemical treatment of the recirculating cooling water
- Intimate contacting of air-water in the cooling device.

Makeup water to recirculating cooling systems replenishes water loss due to evaporation, entrainment (or drift) and blowdown. Makeup water brings soluble chemical species such as sodium (Na^+), potassium (K^+), calcium (Ca^{++}), magnesium (Mg^{++}), chloride (Cl^-), nitrate (NO_3^-), sulfate ($SO_4^=$), and carbon dioxide (HCO_3^- and $CO_3^=$) into the system. The degree of concentration of these species is governed by the operating characteristics of the cooling system, such as blowdown, drift and evaporation rates. Soluble constituents in makeup water become concentrated to levels typically ranging from 1,500 to 10,000 mg/ℓ (ppm) before being removed in the blowdown stream (MA-230). The chemical species contributing to the salinity of the blowdown is primarily determined by makeup water composition.

Chemical treatment is commonly practiced in recirculating cooling systems to control corrosion, scale, biological fouling, and solids deposition. Table 5.1-1 summarizes some of the treatment methods employed and also presents their impact on the quality of the blowdown stream.

The intimate contact between air and water in the cooling device enables particulate matter and soluble gases to be scrubbed from the air contacted. Airborne solids captured by the cooling water significantly contribute to the solids that accumulate in the cooling system. It is estimated that up to 80% of the suspended solids in recirculating systems originally come into the system as airborne particulates (GL-028). Upon dissolution, water soluble particulates will increase the concentration of dissolved species of the circulating water as well. Soluble gases give rise to anionic species in the cooling water. For example, carbon dioxide (CO_2), nitrogen oxides (NO_x), sulfur oxides (SO_x) yield carbonates ($CO_3^=$ and HCO_3^-), nitrates (NO_3^-), and sulfates ($SO_4^=$) in the cooling water, respectively, when these gases are scrubbed from the air.

Leaching of preservatives from treated wood cooling towers constitutes an additional source of potentially hazardous components in cooling water blowdown. Preservatives commonly used include acid copper chromate (ACC), chromated copper arsenate (CCA), creosote and pentachlorophenol. The extent of leaching is currently under investigation.

Additional potential contaminants which may apply in some cases include insecticides and herbicides from agricultural runoff, or phenolic compounds from vegetation decay, most of which are considered toxic. Chlorine addition to control

TABLE 5.1-1. CHEMICAL TREATMENT SUMMARY FOR RECIRCULATING COOLING SYSTEMS

Treatment Objective	Chemical Additive	Typical Additive Concentrations in Blowdown	Comments	Reference
Corrosion Inhibition	Chromate	10-50 mg/ℓ as CrO_4	Chromate treatment has been the traditional corrosion inhibitor system. Since chromate has been found to be highly toxic to aquatic life, water treatment vendors are now offering alternative corrosion inhibitor treatments which employ various combinations of chromate, zinc, phosphate, silicate, and organic additives These alternative treatments are designed to either minimize the chromate concentration that is necessary for corrosion protection or to completely eliminate the need for chromate by substituting other chemical additives	AY-007 GL-028
	Zinc	8-35 mg/ℓ as Zn		
	Phosphate	15-60 mg/ℓ as PO_4		
	Silicates	----		
	Proprietary Organics	3-10 mg/ℓ as organic		
Scale Control	Acid Treatment	Cooling water pH is maintained between 6.5 and 8.0.	Scale control allows recirculating cooling systems to operate at higher concentration factors without the formation of scale on condenser heat transfer surfaces Acid treatment, polyphosphates, and chelating agents maintain the solubility of the common scaling salts (i.e., $CaCO_3$, $CaSO_4$, etc) below the scaling limit (the point at which they will precipitate from solution). Polyelectrolyte antiprecipitants allow supersaturation of the cooling water with respect to scaling salts without precipitation of these salts occurring. Dispersants do not inhibit scale precipitation, but prevent precipitated salts from settling and adhering to heat transfer surfaces	AY-007 MA-230
	Inorganic Polyphosphates	2-5 mg/ℓ		
	Chelating Agents	----		
	Polyelectrolyte Antiprecipitants	1-2 mg/ℓ		
	Organic/Polymer Dispersants	20-50 mg/ℓ		
Biological Fouling (algae, slimes, fungi) Control	Chlorine	≤ 0.5 mg/ℓ residual Cl_2	Biocides used to control biological fouling are either the oxidizing or non-oxidizing types Oxidizing biocides (chlorine and hypochlorite) have been discussed for once-through cooling systems in the "Once-Through Cooling Water" section. These biocides are used in recirculating cooling systems in a fashion similar to that described for once-through systems Non-oxidizing biocides (chlorophenates, thiocyanates, organic sulfur compounds, etc) are employed when other chemical additives such as organic corrosion inhibitors, scale control agents, or solids control agents are destroyed by the conventional oxidizing biocides.	EN-127 AY-007 MA-230
	Hypochlorite	----		
	Chlorophenates			
	Thiocyanates			
	Organic Sulfur Compounds	~ 30 mg/ℓ residual concentrations		
Suspended Solids Dispersion	Tannins	----	Chemical dispersants maintain suspended solids from settling and adhering to heat transfer surfaces.	AY-007 DO-048
	Lignins	----		
	Proprietary Organics/Polymers	20-50 mg/ℓ		
	Polyelectrolytes/Nonionic Polymers	1-2 mg/ℓ		

biological fouling can result in chlorination of these or other hydrocarbons entering with the makeup and result in highly undesirable reaction products.

Ash Handling

Ash is a solid by-product of coal combustion and appears in a power plant boiler in two distinct forms: bottom ash and fly ash. Bottom ash must be removed from the boiler in order to maintain system operability. Fly ash is normally collected in flue gas cleaning equipment. The conveyance of both bottom ash and fly ash to their ultimate points of disposal constitutes ash handling.

Ash handling systems employ either pneumatic or hydraulic mechanisms for ash transportation. This section addresses only the hydraulic (or wet sluicing) systems. Fly ash collected dry has a market in some locations, and therefore, would not only eliminate one major wastewater source, but would be a salable by-product. The usual method of operation, however, is to wet sluice the ash to a pond for settling, or truck the ash dry to some other location for off-site disposal. Wet sluicing of ash is a major source of wastewaters.

Coal-fired generating stations require formal ash handling facilities due to the quantity of ash produced during coal combustion. The ash content of U.S. coals ranges from 6 to 20 wt %. The average value is approximately 11 wt % (EN-127). The distribution between bottom ash and fly ash is greatly influenced by boiler furnace design and operating mode. The ash distribution can affect the water balance for a hydraulic ash handling system. The chemical differences between fly ash and bottom ash can also affect sluicing water quality.

Bottom ash generally forms as a fused, clinker-type material and is removed by wet sluicing. Hydraulic design considerations dictate the minimum sluice water requirements as 10-20 kg (tons) per kg (tons) of bottom ash transported. In actual practice, as high as 165 kg (tons) of water per kg (ton) of bottom ash are used depending on such factors as plant design, location, and operating circumstances (AY-007). Bottom ash has excellent settling characteristics; therefore, sluice water will be relatively free of suspended solids if adequate residence time is supplied for sedimentation. The chemical composition of sluice will not significantly change from sluice influent due to the chemically inert nature of bottom ash with water (SC-267, AS-054).

Fly ash is collected in the dry form by cyclones, fabric filters, dry electrostatic precipitators, etc., and in a water slurry by wet scrubbers, wet electrostatic precipitators, etc. Fly ash collected in either the wet or dry form is commonly sluiced to ash ponds for sedimentation of the suspended fly ash solids. Sluice water in the pond may be 1) discharged as a waste effluent, 2) recycled for additional ash sluicing, or 3) evaporated where meteorological conditions are favorable. The minimum sluice water quantities are set by hydraulic design considerations. For fly ash, the minimum sluice water requirement ranges from 10 to 20 kg (tons) per kg (ton) of fly ash transported. However, as with bottom ash sluicing, the sluice water requirement may be as high as 165 kg (tons) per kg (ton) of ash (AY-007). Flows range from 4.5 to 150 m^3 per day/MW (1200 to 40,000 gpd/MW), while a typical rate for coal is 40 m^3 per day/MW (10,000 gpd/MW).

Although fly ash has somewhat poorer settling characteristics than does bottom ash, low turbidities are observed in sluice effluents if adequate retention time in the ash pond is provided. The residence time necessary varies for different fly ash. It is determined by the settling velocity of the particles and the rate of flow through the ash pond. The particle settling velocity depends on particle size. Most fly ash settles compactly in a pond volume of about 0.6 m^3/metric ton dry fly ash solids (20 ft^3/ton solids) (IF-001). Fly ash contains a broad spectrum of soluble inorganic salts which give rise to sodium, potassium, calcium, magnesium, chlorides, sulfates, etc., in solution. The level of these dissolved solids in solution may range from a few hundred to many thousand mg/ℓ (ppm). In addition, varying concentrations of approximately 30 different trace elements have been detected in both bottom ash and fly ash sluice water (AS-054). Table 5.1-2 presents ash-pond effluent analyses for a large coal-fired plant where separate bottom ash and fly ash ponds are employed.

Sluice water pH is also affected by soluble chemical species in fly ash. Fly ash from pulverized coal burning units contain alkaline species such as oxides of sodium, potassium, calcium, and magnesium (Na_2O, K_2O, CaO, and MgO). Dissolution of these salts can increase sluice water pH to levels on the order of 10.0. On the other hand, fly ash from cyclone furnaces can yield sluice water pH as low as 5.5 due to adsorption of acidic species such as SO_2, SO_3, HCl, etc., on fly ash surfaces (AS-054). The type of coal also influences sluice pH. Western coals generally produce ashes with high lime content, while eastern coal ashes contain lower levels of alkaline species and, in some cases, higher chloride levels, resulting in acidic ash pond effluents.

TABLE 5.1-2. CHARACTERISTICS OF ONCE-THROUGH ASH POND DISCHARGES

Parameter	Flyash Pond			Bottom Ash Pond		
	Min	Ave	Max	Min	Ave	Max
Flow m^3/s (gpm)	0.20 (3,100)	0.39 (6,200)	0.56 (8,800)	0.28 (4,500)	1.00 (16,000)	1.45 (23,000)
Total alkalinity (as $CaCO_3$)	-	-	-	30	85	160
Phen. alkalinity (as $CaCO_3$)	0	0	0	0	0	0
Conductivity (µmhos/cm)	615	810	1125	210	322	910
Total hardness (as $CaCO_3$)	185	260.5	520	76	141.5	394
pH	3.6	4.4	6.3	4.1	7.2	7.9
Dissolved solids	141	508	820	69	167	404
Suspended solids	2	62.5	256	5	60	657
Aluminum	3.6	7.19	8.8	0.5	3.49	8.0
Ammonia (as N)	0.02	0.43	1.4	0.04	0.12	0.34
Arsenic	<0.005	0.010	0.023	0.002	0.006	0.015
Barium	0.2	0.25	0.4	<0.10	0.15	0.30
Beryllium	<0.01	0.011	0.02	<0.01	<0.01	<0.01
Cadmium	0.023	0.037	0.052	<0.001	0.0011	0.002
Calcium	94	136	180	23	40.12	67
Chloride	5	7.12	14	5	8.38	15
Chromium	0.012	0.067	0.17	<0.005	0.009	0.023
Copper	0.16	0.31	0.45	<0.01	0.065	0.14
Cyanide	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
Iron	0.33	1.44	6.6	1.7	5.29	11
Lead	<0.01	0.058	0.2	<0.01	0.016	0.031
Magnesium	9.4	13.99	20	0.3	5.85	9.3
Manganese	0.29	0.48	0.63	0.07	0.16	0.26
Mercury	<0.0002	0.0003	0.0006	<0.0002	0.0007	0.0026
Nickel	0.06	1.1	0.13	0.05	<0.059	0.12
Total Phosphate (as P)	<0.01	0.021	0.06	<0.01	0.081	0.23
Selenium	<0.001	0.0019	0.004	<0.001	0.002	0.004
Silica	10	12.57	15	6.1	7.4	8.6
Silver	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01
Sulfate	240	357.5	440	41	48.75	80
Zinc	1.1	1.51	2.7	0.02	0.09	0.16

Units are mg/ℓ (ppm) unless otherwise indicated

SOURCE: RI-160

Water Conditioning

Some degree of water treating is practiced in all power plants to remove suspended solids and/or dissolved mineral salts. The basic water treating processes used by utilities include:

• sedimentation	suspended solids removal
• filtration	
• lime/lime-soda softening	dissolved solids removal
• ion exchange	
• evaporation	

Water conditioning schemes employ these basic processes singly or in multiple combinations. The subsections below characterize the wastewater expected from each basic process.

Sedimentation

Sedimentation is a solid-liquid separation process in which suspended solids in dilute concentrations are separated from water and concentrated by gravity settling. Turbid water is charged to a clarifier and the suspended solids settle and concentrate at the vessel's bottom. The clarifier underflow constitutes the waste stream for the process. It ranges from 1.0 to 4.0% of the total water charge. The solids content of the underflow rarely exceeds 5.0 wt% (AY-007) and typically is on the order of 2.5 wt% (EN-127).

The liquid phase composition of the clarifier underflow is influenced by coagulating the flocculating agents that may be added to the process in low concentrations. Table 5.1-3 characterizes the types of coagulants and flocculants that are commonly used as well as their normal concentrations in clarification equipment. Other chemical characteristics of the clarifier underflow's liquid phase are essentially equivalent to that of the influent water to the process.

Filtration

Filtration, another solid-liquid separation process, is used extensively to remove suspended solids from water used in the electric utility industry. During the service cycle, turbid water is passed through the granular filter media of the pressure filter which collects solid impurities in the water. These solids accumulate in the media until they are removed by periodic backwashing of the filter.

The wastewater generated by the filtration process comes as a result of filter backwashing. This process involves passing a large surge of water in an upward direction through the filter media. Backwashing expands the filter bed and provides sufficient turbulence for solids to be freed from the media and swept from the filter. The backwash quantity is highly variable; it can be as high as 6.0% of the water treated. Backwash effluent quality has a high suspended solids content which varies with the solids-holding filter capacity, the backwash efficiency, and the backwash water quantity. Filter backwash from plants producing an alum sludge typically contains 40 to 100 mg/ℓ (ppm) suspended solids. Other water quality parameters such as TDS, pH, hardness, etc. are roughly equivalent to the backwash influent, and thus, are largely determined by the choice of backwash water.

TABLE 5.1-3. COAGULATING AND FLOCCULATING AGENT CHARACTERISTICS

Coagulant/Flocculant	Purpose	Normal Dosage (mg/ℓ)
Alum $Al_2(SO_4)_3 \cdot 14\ H_2O$	Main coagulant To assist coagulation with aluminate	5-50 2-20
Aluminate $Na_2Al_2O_4$	Main coagulant To assist coagulation with alum	5-15 ∿2 (0.1 to 0.05 of alum dosage)
Ferric Chloride $FeCl_3 \cdot 6H_2O$	Main coagulant	5-50
Copperas $FeSO_4 \cdot 7H_2O$	Main coagulant	5-50
Weighting Agents (bentonite, kaolin, montmorillonite)	Coagulant Aid	----
Adsorbents (powdered carbon, activated alumina)	Coagulant Aid	----
Polyelectrolytes (inorganic activated silica and organic polymers)	Coagulant Aid	≤ 2

SOURCES: AY-007, TE-111

Lime/Lime-Soda Softening

Lime/lime-soda softening is employed to treat the hardness and alkalinity content of water. Calcium and magnesium are precipitated as calcium carbonate ($CaCO_3$) and magnesium hydroxide [$Mg(OH)_2$], respectively. Hydrated lime [$Ca(OH)_2$] and/or soda ash (Na_2CO_3) are added to the process to drive chemical precipitation. Precipitates of $CaCO_3$ and $Mg(OH)_2$ are separated from the treated water by sedimentation.

The wastewater problem for the softening process arises from removal of precipitated salts. Many similarities between softener wastes and sedimentation wastes (described in the Sedimentation Section) exist. The solids concentration of the softener waste is typically 2.5 wt % but may be as high as 5.0 wt %. The waste quantity varies with influent water hardness and process softening efficiency.

The liquid composition of the waste is similar to the softened water effluent. It contains coagulants and flocculants to enhance sedimentation efficiency. Total hardness is typically on the order of 50 mg/ℓ (ppm) as $CaCO_3$ (ST-135). Wastewater pH is approximately 10.0 due to the alkaline reagents (soda ash and/or lime) added to the process.

Ion Exchange

Ion exchange is used to provide softening and demineralization capability in power plant water treatment systems. Water softening by ion exchange can be used for condenser cooling systems, low pressure boilers, and other evaporative processes which have a water concentrating effect. Demineralization is used exclusively as a water treatment for high pressure boilers.

Ion exchange resins are the heart of the process and they are classified as either the cationic or anionic type depending upon their ion affinity. As the exchange resins reach their ion load capacity, they are removed from service and are regenerated. Regeneration is accomplished in three distinct steps:

- Backwashing for resin bed expansion and solids removal
- Elution of ions from the resin with a regenerant solution
- Rinsing excess regenerant and eluted ions from the resin bed.

All three regeneration steps contribute to wastewater generated in the ion exchange process.

Backwashing loosens and expands the resin bed and removes suspended solids that accumulate on the resin during service flow. This process is accomplished by an upward flow of water through the resin bed. The backwash flow rate is adjusted to expand the bed volume to 80-100% greater than its settled depth; this flow is maintained for a duration of 10 to 15 minutes. Flow rates of 3.4 - 4.1 $m^3/sec/m^2$ (5-6 gallons per minute per ft^2 of bed cross-sectional area) are typical (EN-127). Raw influent to the process is generally employed as backwash water. The chemical quality of the backwash water will not be significantly altered as it passes through the ion exchange bed.

After backwashing, the ion exchange resins are treated with regenerant solutions to rejuvenate ion exchange capability. The regenerant solutions elute ions from the ion exchange resins and replace them with species from the regenerant solution. Hydrogen cycle cation exchange resins are regenerated with acid solutions while anion exchange resins are regenerated with alkalis. The spent regenerant solutions represent the aqueous waste associated with this step of the regeneration sequence.

Spent regenerant solutions contain ions that are eluted from the ion exchange material plus the excess regenerant that is not consumed during regeneration. The eluted ions represent the chemical species that were removed from water during the service cycle of the process. For example, regeneration of a hydrogen cycle cation exchange resin will elute such species as sodium, potassium, calcium, magnesium, etc. Similarly, ions eluted from anion exchange units will include chloride, nitrate, bicarbonate, sulfate, carbonate, etc. The distribution of these species in spent regenerant solutions vary with influent water to the ion exchange process.

The excess regenerant required for ion elution varies with the ion exchange resins employed. Table 5.1-4 presents a summary of ion exchange material types and the regenerant requirements of each. With the exception of sodium cycle ion exchange, excess regenerant creates an effluent of extreme pH. Spent regenerant from cation exchange units is acidic (low pH). Alkaline (high pH) regenerants are characteristic of anion exchange units.

The final step of regeneration is rinsing spent regenerant solution from the ion exchange bed. The rinse water quantity varies with each resin type, but it is typically 8.0 m^3 of water per m^3 of resin for cationic resins and 10 m^3 of water per m^3 of resin for anionic resins (EN-127). Rinse

TABLE 5.1-4. ION EXCHANGE MATERIAL TYPES AND REGENERANT REQUIREMENT

Ion Exchange Material	Description of Operation	Regenerant Solution	Theoretical Amount
Cation Exchange Sodium Cycle	Sodium cycle ion exchange is used as a water softening process. Calcium, magnesium, and other divalent cations are exchanged for more soluble sodium cations, i.e., $2R_c\text{-Na} + Ca^{++} \rightleftarrows (R_c)_2\text{-Ca} + 2\ Na^+$ $2R_c\text{-Na} + Mg^{++} \rightleftarrows (R_c)_2\text{-Mg} + 2\ Na^+$	10% brine (NaCl) solution or some other solution with a relatively high sodium content such as sea water	————
Hydrogen Cycle Weak Acid	Weak acid ion exchange removes cations from water in quantities equivalent to the total alkalinity present in the water, i.e., $2R_c\text{-H} + Ca(HCO_3)_2 \rightleftarrows (R_c)_2\text{-Ca} + 2H_2CO_3$	H_2SO_4 or HCl solutions with acid strengths as low as 0.5%	110 -120%
Strong Acid	Strong acid ion exchange removes cations of all soluble salts in water, i.e.,	H_2SO_4 or HCl solutions with acid strengths renging from 2.0 - 6.0%	200 - 400%
Anion Exchange Weak Base	Weak base ion exchange removes anions of all strong mineral acids (H_2SO_4, HCl, HNO_3, etc:), i.e., $2R_A\text{-OH} + H_2SO_4 \rightleftarrows (R_A)_2\text{-}SO_4 + 2HOH$	NaOH, NH_4OH, Na_2CO_3 solutions of variable strength	120 - 140%
Strong Base	Strong base ion exchange removes anions of all soluble salts in water, i.e., $R_A\text{-OH} + H_2CO_3 \rightleftarrows R_A\text{-}HCO_3 + HOH$	NaOH solutions at approximate 4.0% strength.	150 - 300%

Sources: ST-135, DE-079

water quality characteristics range from those of the spent regenerant solution to those of treated water.

Evaporation

Evaporation is a demineralization process sometimes used to treat boiler feedwater whereby raw feedwater is distilled to produce a pure condensate. Feedwater impurities concentrated in the evaporator are removed as a waste blowdown stream.

Evaporator blowdown has a high dissolved solids content. The concentration of the dissolved solids varies with the level of dissolved salts in the influent water and with the degree of concentration in the evaporator. The degree of concentration is limited so that the evaporator operates within the solubility limit of calcium, magnesium, and other scaling salts. Therefore, as the scaling potential of influent water increases, the allowable degree of concentration in the evaporator decreases.

The blowdown quantity necessary to maintain acceptable scaling potentials ranges from 10-40% of the water charged to the process (AY-007). This corresponds to maximum TDS of approximately 3000 mg/ℓ. However, concentrations on the order of 1000-2000 mg/ℓ are more commonly observed (EN-127, AY-007). The distribution of soluble species in the blowdown is similar to the distribution for influent water. However, blowdown pH is typically 9 to 11 due to the thermal decomposition of carbonates in the evaporator (EN-127).

Boiler Blowdown

Power plant boilers are either of the once-through or the drum-type design. Once-through designs are employed exclusively in high pressure, super-critical boilers and have no wastewater streams directly associated with their operation. Thus, they will not be considered further in this section. Drum-type boilers, on the other hand, operate at sub-critical conditions where steam generated in the drum-type units is in equilibrium with liquid boiler water. Boiler water impurities are, therefore, concentrated in the liquid phase as steam is generated in these units. These impurities are ultimately removed in a liquid blowdown stream, the wastewater from this system.

The blowdown from drum-type boilers generally contains soluble inorganic species that occur in natural waters (i.e., Na^+, K^+, Cl^-, $SO_4^=$, etc.); precipitated solids containing the calcium/magnesium cation; soluble and insoluble corrosion products of iron, copper, and other metals; plus a variety of chemical compounds added to the system. Dissolved solids are present in excess of all other boiler water impurities. The concentration of impurities in drum-type boiler blowdown is largely governed by boiler operating conditions, such as pressure. Table 5.1-5 presents recommended limits of total and suspended solids in drum-type boilers as a function of drum pressure.

A number of chemical additives may be present in the boiler blowdown as a result of internal boiler water treatment. Internal treatment is designed to control scale formation, corrosion, pH, and solids deposition in the boiler system. A summary of these internal treatment control practices is presented in Table 5.1-6.

TABLE 5.1-5. RECOMMENDED LIMITS OF TOTAL SOLIDS AND SUSPENDED SOLIDS IN BOILER WATER FOR DRUM BOILERS

Limits Recommended for Total (Dissolved and Suspended) Solids

Drum Pressure MPa	Drum Pressure psi	Total Solids mg/ℓ (ppm)
0 - 2.1	0 - 300	3500
2.1 - 3.1	300 - 450	3000
3.1 - 4.1	450 - 600	2500
4.1 - 5.2	600 - 750	2000
5.2 - 6.2	750 - 900	1500
6.2 - 6.9	900 - 1000	1250
6.9 - 10.3	1000 - 1500	1000
10.3 - 13.8	1500 - 2000	750
>13.8	>2000	15

Limits Recommended for Suspended Solids

Drum Pressure	Below 4.1 MPa (600 psi)	4.1-6.9 MPa (600-1000 psi)	6.9-13.8 MPa (1000-2000 psi)	Over 13.8 MPa (2000 psi)
Total solids, mg/ℓ	*	*	0.15	0.05
Total hardness as mg/ℓ $CaCO_3$	0	0	0	0
Iron, mg/ℓ	0.1	0.05	0.01	0.01
Copper, mg/ℓ	0.05	0.03	0.005	0.002
Oxygen, mg/ℓ	0.007	0.007	0.007	0.007
pH	8.0-9.5	8.0-9.5	8.5-9.5	8.5-9.5
Organic	0	0	0	0

*No value reported.

Source: BA-185

TABLE 5.1-6. CHEMICAL ADDITIVES COMMONLY ASSOCIATED WITH INTERNAL BOILER TREATMENT

Control Objective	Candidate Chemical Additives	Residual Concentration in Boiler Water	Reference
Scale	di- and tri-sodium phosphates	3-60 mg/ℓ as PO_4	BA-185
	Ethylene diaminetetracetic acid (EDTA)	20-100 mg/ℓ	EN-127, AY-007, BL-036
	Nitrilotriacetic acid (NTA)	10-60 mg/ℓ	EN-127, AY-007, BL-036
	Alginates	up to 50-100 mg/ℓ	AY-007, BL-036
	Polyacrylates	up to 50-100 mg/ℓ	AY-007, BL-036
	Polymethacrylates	up to 50-100 mg/ℓ	AY-007, BL-036
Corrosion	Sodium sulfite and catalyzed sodium sulfite	less than 200 mg/ℓ	MA-230, BL-036
	Hydrazine	5-45 mg/ℓ	AY-007
	Morpholine	5-45 mg/ℓ	AY-007
pH	Sodium hydroxide	Added to adjust boiler water pH to the desired level, typically 8.0 - 11.0.	EN-127, AY-007, BA-185, BL-036
	Sodium carbonate		
	Ammonia		
	Morpholine		
	Hydrazine		
Solids Deposition	Starch	20-50 mg/ℓ	AY-007
	Alginates	20-50 mg/ℓ	AY-007
	Polyacrylamides	20-50 mg/ℓ	AY-007
	Polyacrylates	20-50 mg/ℓ	AY-007
	Polymethacrylates	20-50 mg/ℓ	AY-007
	Tannins	$\leq$200 mg/ℓ	AY-007
	Lignin derivatives	$\leq$200 mg/ℓ	AY-007

The blowdown quantity of modern, high-pressure boiler ranges from effectively zero to an upper limit of 2.0% of the steam generation rate. The blowdown rate is typically 0.1% of the steam generation rate (AY-007). Much higher blowdown rates, typically 10%, are associated with lower pressure steam generating systems where makeup is not demineralized. Boiler blowdown may be performed in either an intermittent or continuous fashion.

Coal Pile Runoff

Coal-fired power stations maintain reserve fuel on the plant premises in active and/or inactive coal storage piles. Active coal storage is open and is exposed to all ambient conditions. Inactive coal piles are commonly sealed with a tar spray or some other impervious covering which provides protection from the weather. Runoff from active coal storage piles is of primary concern in this section.

Precipitation runoff from active coal storage piles presents a potential problem of stream and ground water pollution. This runoff commonly exhibits extreme pH and contains soluble chemical species and suspended solids. The primary cause of runoff contamination is a reaction mechanism similar to the one that produces acid mine drainage. Inorganic sulfur in the coal reacts with moisture and oxygen in air to produce sulfuric acid.

When rainwater seeps into the coal pile, sulfuric acid is leached from the coal. The pH of the runoff effluent can be as low as 2-3 units. The acidic nature of this water drives the dissolution of inorganic salts that are present in the coal. In addition to a high sulfate anion concentration, the runoff contains high concentrations of cations such as iron, aluminum and manganese. Traces of cadmium, beryllium, nickel, chromium, vanadium, zinc, and copper have also been reported. Coal fines and other insoluble material appear in the runoff as suspended solids.

Table 5.1-7 presents plant data for coal pile runoffs. Coal type has a great influence on runoff characteristics. For example, some coals such as are burned at Plant 5305 have sufficient alkalinity to neutralize all of the sulfuric acid formed. The resulting effluent pH in such cases ranges from 6.5 to 7.5. The higher pH range decreases the solubility of many inorganic salts, thus affecting runoff effluent quality. The runoff at Plants 1729, 3626, 0107, on the other hand, is very acidic because of the high-sulfur fuel burned. Other factors causing variations in the effluent quality besides coal type are coal pile history and runoff flow rate.

The quantity of runoff effluent is a strong function of coal pile area and local meteorological conditions. Coal pile area is primarily determined by generating station size. Power plants store from 600 to 1,800 cubic meters (0.5 to 1.5 acre-feet) of coal for each MW of generating capacity. The storage piles are typically 8 to 12 meters (25 to 40 feet) in height (AY-007). This corresponds to a coal storage area of 50 to 250 square meters (0.013 to 0.060 acres) for each MW of capacity, depending on pile height. An annual precipitation rate of one meter (40 inches), for example, will result in an annual runoff of between 50 and 225 cubic meters (13,000 to 60,000 gallons) per MW of generating capacity. The typical runoff rate is 76,000 - 95,000 cubic meters (20 to 25 million gallons) per year at most coal-fired generating stations (EN-127, AY-007).

TABLE 5.1-7. PLANT DATA RELATING TO WATER QUALITY PARAMETERS FOR COAL PILE RUNOFF

Plant Code	3402	3401	3936	1325	1726	1729	3626	3107	5305	5305	5305
* Alkalinity (mg/ℓ)	5	0	0	-	32	-	-	0	21.36	14.32	36.41
BOD (mg/ℓ)	0	0	10	-	3	-	-	-	-	-	-
COD (mg/ℓ)	1080	1080	306	35	1099	-	-	-	-	-	-
TS	1330	1330	9999	6000	3549	-	-	45000	-	-	-
TDS	720	720	7743	5800	247	-	28970	44050	-	-	-
TSS	610	610	22	200	3302	-	100	950	-	-	-
Ammonia	0	0	1.77	1.35	0.35	-	-	-	-	-	-
Nitrate	0.3	0.3	1.9	1.8	2.25	-	-	-	-	-	-
Phosphorous	-	-	1.2	-	0.23	-	-	-	-	-	-
Turbidity	505	505	-	-	-	-	-	-	8.37	2.77	6.13
Acidity	-	-	-	-	-	-	21700	27810	8.68	10.25	3.34
Total Hardness	130	130	1109	1850	-	-	-	-	-	-	-
Sulfate	525	525	5731	361	133	6837	19000	21920	-	-	-
Chloride	3.6	3.6	481	-	23	-	-	-	-	-	-
Aluminum	-	-	-	-	-	-	1200	325	-	-	-
Chromium	0	0	0.37	0.05	-	-	15.7	0.3	-	-	-
Copper (mg/ℓ)	1.6	1.6	-	-	-	-	1.3	3.4	-	-	-
Iron (mg/ℓ)	0.168	0.168	-	0.06	-	0.368	4700	93000	1.0	1.05	0.9
Magnesium (mg/ℓ)	-		89	174	-	-	-	-	-	-	-
Zinc (mg/ℓ)	1.6	1.6	2.43	0.0006	0.08	-	12.3	23	-	-	-
Sodium (mg/ℓ)	1260	1260	160	-	-	-	-		-	-	-
pH	2.8	2.8	3	4.4	7.8	2.7	2.1	2.8	6.7	6.6	6.6

* All concentrations (except pH) are expressed in mg/ℓ.

Source: EN-127

Chemical Cleaning

Operational cleaning of heat transfer surfaces is designed to remove scale and corrosion products that accumulate on the boiler's steam-side and the water-side of the steam condenser. The frequency at which chemical cleaning is needed varies from plant to plant. For example, the mean time between boiler chemical cleanings is approximately 36 months. However, plant data indicate extreme variations in frequency ranging from once in 7 months to once in 100 months (EN-127).

The active reagents in cleaning solutions are acidic or alkaline in nature depending primarily upon the deposits they are to attack. Ninety percent of all cleaning operations employ acidic formulations that attack all forms of alkaline scale (i.e., $CaCO_3$, $Mg(OH)_2$, etc.), silica scale, and corrosion deposits containing iron. The majority of these formulations contain hydrochloric acid in solution strengths ranging from 5.0 to 7.5% (AY-007). Other acid solutions contain the following constituents which are present alone or in various combinations (EN-127, AY-007, BA-185, BE-162):

Inorganic Acids

- Hydrochloric (HCl)
- Sulfuric (H_2SO_4)
- Sulfamic (NH_2SO_3H)
- Phosphoric (H_3PO_4)
- Nitric (HNO_3)
- Hydrofluoric (HF)

Organic Acids

- Citric [$HOC(CH_2CO_2H)_2CO_2H$]
- Formic (HCO_2H)
- Hydroxyacetic ($HOCH_2CO_2H$)

Representative data for three separate cleaning operations are presented in Tables 5.1-8 through 5.1-10. The data are for cleaning the steam-side of a high pressure, once-through boiler and a low pressure, drum boiler plus the water-side of a main condenser, respectively. Alkaline and acidic solutions are shown for both cleanings while only an acidic solution is shown for condenser cleaning.

Spent chemical cleaning solutions usually have extreme pH, high dissolved solids concentrations, and significant oxygen demands (BOD and/or COD). The pH of spent solutions ranges from 2.5 to 11.0 depending on whether acidic or alkaline cleaning reagents are employed. The dissolved solids include sodium, hardness, heavy metals, chloride, bromide, and fluoride. Tables A-8 through A-10 report only iron and copper concentrations for heavy metals. However, additional metal constituents may include nickel, zinc, and aluminum. Heavy metals for combined boiler cleaning wastes follow the general concentration trend of (EN-127):

iron > copper > nickel > zinc > aluminum

The quantity of cleaning wastes varies directly with liquid holding volume of equipment to be cleaned (BA-185). For example, Tables A-8 and A-9 show spent alkaline and acid cleaning solution quantities as equal to the tube-side boiler volume. Rinse solution quantities are generally one or two times this volume.

TABLE 5.1-8. OPERATIONAL CLEANING OF A HIGH PRESSURE, ONCE-THROUGH BOILER(1)

	Stage 1	Stage 1 Rinse	Stage 2	Stage 2 Rinse
pH	11.0	9.0	2.5-3.0	6.0-7.0
Acidity, ppm total hot, as $CaCO_3$	-	-	27,000	low
Alkalinity, ppm as $CaCO_3$	90,000	9,000	-	-
NH_3, (%)	1	0.1	(2)	(2)
Fe^{++}, ppm	-	-	60,000	600
Cu	720+	75	-	-
BOD	-	-	high	high
Suspended Solids, ppm	500	5	100	5
Volume, gal	30,000	60,000	30,000	60,000
m^3	110	220	110	220
Temperature,				
as drained, °F	100	100	200	200
°C	40	40	90	90

(1) The use of ammonium salts of organic acids raises this number.

(2) Capacity-20,000 gallons; solvent system Stagel, ammonium persulfate solvent; Stage 2, inhibited 3% organic acid; deposit inventory, 200 lbs copper as Cu and 1500 lbs iron as Fe_2O_3.

Source: BE-162

TABLE 5.1-9. OPERATIONAL CLEANING OF A LOW PRESSURE, DRUM BOILER(1)

	Stage 1	Stage 1 Rinses	Stage 2	Stage 2 Rinses
pH	11.0	9.0	<1.0	7.0-9.0
Acidity, ppm total hot as $CaCO_3$	-	-	70,000	-
Alkalinity, ppm as $CaCO_3$	90,000	9,000	-	1,000
NH_3(%)	1	0.1	0.01	-
Chloride, ppm	-	-	68,000	<6,800
Fe^{++}, ppm	-	-	6,000	600
Cu, ppm	720+	75	75	-
Bromide, ppm	1,500	150	15	-
Suspended Solids, ppm	500	5	500	100
Volume, gal	30,000	60,000	30,000	60,000
m^3	110	220	110	220
Temperature, as drained, °F	150	150	150	125
°C	65	65	65	52

(1)Capacity, 30,000 gallons; solvent system Stage 1, bromate solvent; solvent system Stage 2, inhibited 5% hydrochloric acid and copper complexing agent; deposit inventory, 200 lbs copper as Cu and 1,500 lbs iron as Fe_2O_3.

Source: BE-162

TABLE 5.1-10. OPERATIONAL CLEANING, MAIN CONDENSER WATER-SIDE[1]

	Solvent	Rinses
pH	2.0	7.0-9.0
Acidity, hot, ppm $CaCO_3$	30,000	-
Alkalinity, ppm $CaCO_3$	-	1,000
Ca, Mg, ppm	5,000	50
Fe, ppm total	11,000	100
Cu, dissolved, ppm	1,000	50
Suspended solids, ppm	2,000+	200
Temperature, as drained, °F	140	120
°C	60	49

[1] Capacity - 1000 gallons; solvent system, 10% sulfamic acid, inhibited, 1% NaCl; Rinse - Na_2CO_3 + sodium phosphates; depository inventory, 100 lbs Ca and Mg salts, 100 lbs Fe_2O_3, 10 lbs copper.

Source: BE-162

Washing Operations

Floor and equipment washing operations produce effluents containing suspended solids, detergent constituents, oily wastes, and a broad spectrum of soluble inorganic species. Water quality parameters for washing operations are highly dependent on the specific area or equipment item washed . Two major washing operations at fossil-fueled power plants concern removal of deposits consisting of fuel ash, soot, combustion additives, etc., from the boiler fireside and air preheater.

Boiler firesides are commonly washed by spraying high pressure water against boiler tubes while they are still hot. In some cases alkaline wash water is used. Waste effluents from this washing operation contain an assortment of dissolved and suspended solids. Wash frequencies, waste volumes, and effluent water quality data for two generating stations are presented in Table 5.1-11. It should be noted that acid wastes are common for boilers fired with high sulfur fuels. Sulfur oxides adsorb onto fireside deposits and impart low pH and a high sulfate content to the waste effluent (BA-185). These deposits can also be a source of iron, nickel, chromium, vanadium, and zinc depending on the fuel type and fuel additives.

Air preheaters employed in power stations are either the tubular or regenerative types. Both are periodically washed to remove deposits that accumulate. The frequency of washing is typically once per month; however, frequency variations ranging from 4 to 180 washings per year are reported (EN-127). Many air preheater designs are sectionalized so that heat transfer areas may be isolated and washed without

TABLE 5.1-11. DATA FOR BOILER FIRE SIDE WASHING OPERATIONS; INCREASE IN POLLUTANT QUANTITY PER WASHING CYCLE

Plant Code	3410	3411
Cleaning Frequency cycles/yr	2	3
Batch Volume, m^3	2626	90.8
1000 gal	720	24
Alkalinity, lb	-240	-5.99
kg	-109	-2.72
COD, lb	1134	19
kg	515	8.63
Total Solids, lb	40861	4002
kg	18551	1817
Total Dissolved Solids, lb	35127	3002
kg	15948	1363
Total Suspended Solids, lb	3823	119.09
kg	1736	54.07
Sulfate, lb	11949	299.4
kg	5425	135.9
Chloride, lb	0	18.01
kg	0	8.16
Ammonia, lb	1.49	0.039
kg	0.68	0.018
Nitrate, lb	14.75	0.7
kg	6.7	0.318
Phosphorous, lb	11.1	0.257
kg	5.04	0.117
Hardness, lb	35409	791.41
kg	16076	359.3
Chromium, lb	0.0299	0.998
kg	0.0136	0.453
Copper, lb	-	0.249
kg	-	0.113
Iron, lb	900	30
kg	408.9	13.63
Magnesium, lb	11949	190.35
kg	5425	86.42
Nickel, lb	30.02	-
kg	13.63	-
Sodium, lb	0	9
kg	0	4.09
Zinc, lb	28.72	2
kg	13.042	0.908
BOD, lb	0	0
kg	0	0
Turbidity, JTU	476	98

SOURCE: EN-127

shutdown of the entire unit (BA-185). Higher wash frequencies are expected for air preheaters employing this design feature.

Fossil fuels with significant sulfur content will produce sulfur oxides which adsorb on air preheater deposits. Water washing for these deposits produces an acidic effluent. Alkaline reagents are often added to wash water to neutralize acidity, prevent corrosion of metallic surfaces, and maintain an alkaline pH. Alkaline reagents might include soda ash (Na_2CO_3), caustic soda (NaOH), phosphates and/or detergent.

Preheater wash water contains high solids content (both suspended and dissolved). The solids primarily include sulfates, hardness, and heavy metals, including large quantities of copper, iron, nickel, and chromium. The levels of metals in preheater deposits are usually much higher than those in fireside deposits. Air preheater wastes may also be a source of oily matter and polynuclear hydrocarbons. Data for air preheater wash water discharges are presented in Table 5.1-12.

General Plant Drainage

General plant drainage refers to liquid that accumulates in a floor and yard drains in the process area of a power plant as a result of precipitation runoff.

Plant drainage generally contains a high level of suspended solids consisting of such materials as soil, dust, coal fines, fly ash, etc. that are entrained in the runoff flow. Any significant degree of dissolution of these solids will also add to the dissolved salts present in the water. The specific characteristics of runoff vary radically from plant to plant and from time to time for a given plant. The

TABLE 5.1-12. DATA FOR AIR PREHEATER WASHING OPERATIONS; INCREASE IN POLLUTANT QUANTITY PER WASHING CYCLES

Plant Code	3409	3410	3411	3413	3414	3415
Cleaning Frequency cycles/yr	12	12	3	5	6	4
Batch Volume, m^3	409	852	1363	265	162.8	378.6
1000 gal	108	225	360	70	43	100
COD, lb	14.4	16.87	14.98	116.7	5.72	9.16
kg	6.54	7.66	6.8	53	2.6	4.16
Total Solids, lb	11951	24964	40528	2616	4768	11257
kg	5426	11334	18400	1188	2165	5111
Total Dissolved Solids, lb	7907	16605	27022	4467	3189	3249
kg	3590	7539	12268	2028	1448	3745
Total Suspended Solids, lb	1975	4008	6603	477.9	785.24	1834
kg	897	1820	2998	217	356.5	833
Sulfate, lb	1066	2231	3601	692	423.8	979
kg	484	1013	1635	314.2	192.4	444.5
Chloride, lb	1.301	0	0	0	-8.96	-14.16
kg	0.3178	0	0	0	-4.07	-6.43
Hardness, lb	3949	8255	13372	476.8	1577	3709
kg	1793	3748	6071	216.5	716	1684
Chromium, lb	1.1	24.25	39.03	0.749	0.458	0.533
kg	0.529	11.01	17.72	0.34	0.208	0.242
Copper, lb	4.434	-	-	2.907	1.788	1.36
kg	2.018	-	-	1.32	0.812	0.848
Iron, lb	1531	3189	5103	3.495	2.13	2.379
kg	695.1	1448	2317	1.587	0.967	1.08
Magnesium, lb	874.45	1850	2986	107.4	352.4	828
kg	397	840	1356	48.76	160	376
Nickel, lb	67.55	140.72	225	28.63	17.93	20.83
kg	30.67	63.89	102.2	13	8.14	9.46
Sodium, lb	1.799	0	0	552	-0.35	1.66
kg	0.818	0	0	251	-0.16	0.757
Zinc, lb	4.43	8.97	14.93	0.283	1.788	2.07
kg	2.011	4.075	6.78	0.1285	0.812	0.942
BOD, lb	3.6	0	0	2.335	1.793	1.668
kg	1.635	0	0	1.06	0.814	0.757
Turbidity, JTU	495	476	497	500	500	498

SOURCE: EN-127

general trend is that highest impurity levels are observed as initial runoff occurs. Subsequent runoff is observed to have lower impurity concentrations.

Process Spills and Leaks

Liquid spills and leaks are commonly associated with overfilling of storage vessels; tank or pipe ruptures; failure of valves, pump seals, etc. Waste effluent characteristics resulting from spills and leaks depend upon the type of fluid that escapes containment. Potential fluids include:

- acid, alkalis, and brine solutions for ion exchange regenerants as well as other water treating chemicals
- fuel oil, transformer oil, and circuit breaker oil
- water used in plant operation such as cooling water.

Miscellaneous Operations

Several miscellaneous sources of wastewater at power plants are listed below:

- laboratory and sampling operations
- auxiliary cooling system(s)
- water intake screen washings
- others

Although the impact of wastewater from these miscellaneous operations is less significant than those discussed in previous sections, these sources nevertheless contribute to the total wastewater problem.

Quantitative data which characterize wastewater from these miscellaneous sources is limited. The only data available are for auxiliary cooling systems, which remove heat from mechanical equipment items such as those listed below:

bearing and/or gland cooling for pumps, fans, and other rotary equipment

air compressor water jackets

generator cooling

Auxiliary cooling systems can be either the once-through or closed-cycle types.

Once-through auxiliary cooling systems do not usually involve chemical treatment, with the exception of chlorination. Thus, water quality of the waste is determined by the influent water to the system. During chlorination of these systems, residual chlorine levels in the effluent are approximately the same as presented for condenser cooling systems, about 0.1 to 0.2 mg/ℓ (ppm). However, the frequency of chlorination is considerably reduced. (See sections entitled Once-Through Cooling Water and Cooling Tower Blowdown.)

The flow through the once-through auxiliary cooling circuit ranges from 0.000032 to 0.0022 m^3/s (0.5 to 35 gpm) per MW of rated generating capacity. The typical flow is approximately 0.00066 m^3/s (10 to 11 gpm) (AY-007). This total flow represents the wastewater stream for once-through auxiliary cooling.

Wastewater from a closed-cycle auxiliary cooling system is a blowdown stream. Water in the closed cooling circuit is treated to control corrosion with inhibitors such as chromates (at levels up to 250 mg/ℓ) and borates or nitrates (at levels ranging from 500 to 2000 mg/ℓ). Water pH is maintained between 9.5 and 10 by addition of caustic soda or soda ash (AY-007). Alternatively, some plants use stream condensate ammonia and hydrazone in this cooling circuit (EN-127). The water recirculation rate is typically 0.0015 to 0.0016 m^3/s (23-25 gpm) per MW of rated generating capacity. Blowdown rates vary from zero to 0.0019 m^3/day (0-5 gpd) (EN-127, AY-007).

5.2 Characterization of Wastewaters from the Lime Wet Scrubbing Process

A description of the Lime Wet Scrubbing Process is given in Section 3.2. The following section presents a closer examination of the water system. A water balance for one of four equivalent scrubbing trains for 500 MW power plant is given. An examination of the water system flows allows the conclusion that there are no wastewater streams in the lime wet scrubbing process in normal operation.

5.2.1 Base Case Water Balance

In calculating an example water balance to illustrate the water flows in a lime wet scrubbing system, Radian chose a base case for ease of comparison. The base case is calculated for one of four equivalent scrubbing systems for a 500 MW power plant burning 3.5 percent sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb). The detailed calculations are presented in Appendix B. Figure 5.2-1 illustrates the water

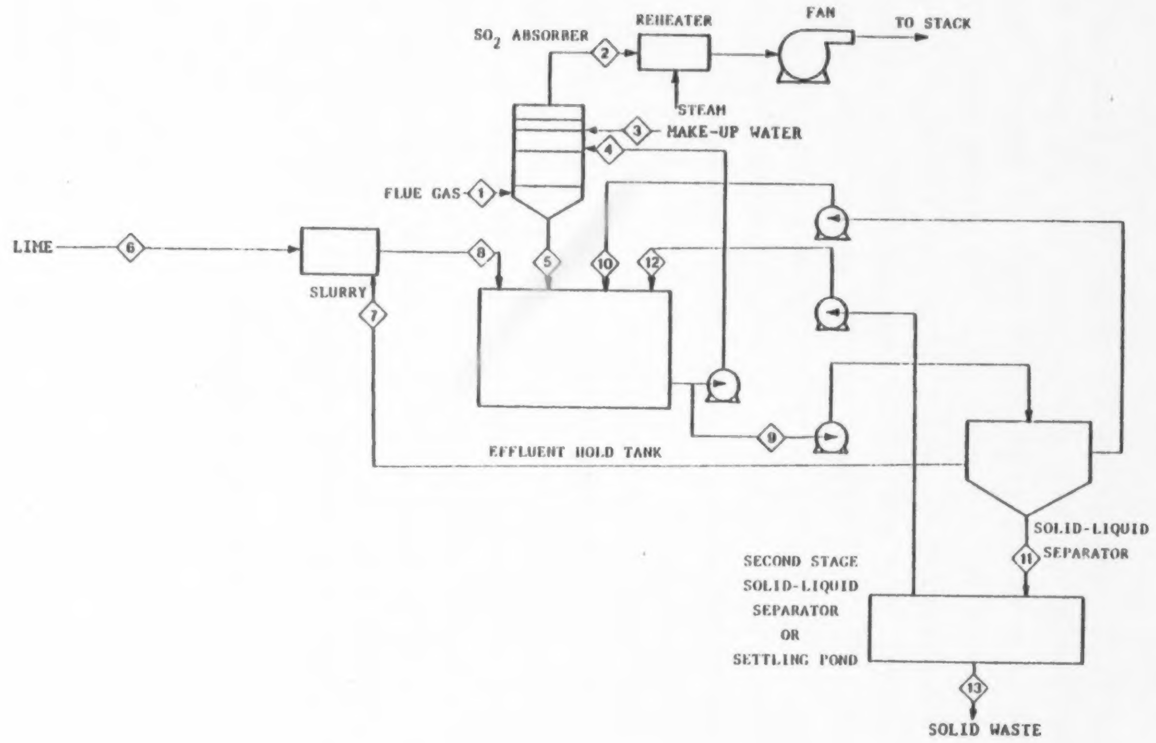

SO$_2$ ABSORBER
REHEATER
FAN
TO STACK
2
STEAM
3
MAKE-UP WATER
4
FLUE GAS
1
LIME
6
SLURRY
7
8
5
10
12
9
EFFLUENT HOLD TANK
SOLID-LIQUID SEPARATOR
11
SECOND STAGE SOLID-LIQUID SEPARATOR OR SETTLING POND
13
SOLID WASTE

Figure 5.2-1. Process flow diagram lime wet scrubbing process.

system. The results of the water balance calculations are presented in Table 5.2.1. Several of the circulating flow rates are characteristic flows rather than design flows, as indicated.

Flue gas first enters an ESP where 99 percent of the fly ash particulates are removed and sluiced to the ash pond. The flue gases, which contain some percentage of water, then enter the absorber where SO_2 is removed. The gases are cooled and saturated by evaporation of water from the circulating lime slurry. The saturated flue gas exits the absorber, is reheated for proper bouyancy, and discharged to the stack. The absorber effluent is sent to a hold tank where calcium sulfite and sulfate crystals are precipitated. A 10-15 weight percent slurry is recirculated to the absorber. A bleed stream is sent to solids dewatering. The supernatant from the solid liquid separator is recycled to the hold tank. The underflow is typically 30-40 percent solids by weight. Either second stage solid-liquid separation or a settling pond is used to recover more of the liquor. This liquor is then recycled to the hold tank. The final sludge is typically 50-60 percent solids in a settling tank, or 60-70 percent solids if a vacuum filter and/or centrifuge is used. Makeup water is therefore required to compensate for loss due to evaporation in the absorber and to occlusion in the solid waste.

5.2.2 Description of the Water System

The water systems in the lime wet scrubbing process can be designed to operate in closed loops, as exemplified by the lime scrubbing systems at Louisville Gas and Electric's Paddy's Run and Pennsylvania Power Company's Bruce Mansfield stations (PE-259). In a closed system, the lime slurry is recirculated between the effluent hold tank and the absorber. A bleed stream is taken for solids dewatering and disposal. The solids are generally removed in a clarifier underflow as a 30-40 weight percent sludge. The clarifier supernatant is recycled

TABLE 5.2-1. WATER BALANCE: LIME WET SCRUBBING PROCESS
(one of four equivalent scrubbing trains)

Stream No. Figure 5.2-1	1	2	3	4*	5*	6	7	8**	9	10	11	12	13
Description	Gas to Absorber	Gas to Air Heater	Make-up Water to Absorber	Recycle Slurry to Absorber	Slurry to Hold Tank	Lime Feed	First Stage S/L Separator to Slurry Feed Tank	Lime Slurry to Hold Tank	Slurry Bleed Stream	Solid/Liquid Separator Overflow to Hold Tank	S/L Separator Underflow to Pond or Filtration	Second Stage S/L Overflow to Hold Tank	Solid Waste
Stream Rate, kg/s (lb/min)	160 (21,000)	179 (22,000)	8.3 (1100)	940 (125,000)	940 (125,000)	0.63 (84)	4.9 (650)	5.6 (740)	15 (2000)	6.3 (830)	3.9 (520)	1.3 (170)	2.6 (350)
Water, kg/s (lb/min)	7.64 (1000)	15.1 (2000)	8.3 (1100)	830 (110,000)	830 (110,000)		4.9 (650)	4.9 (650)	14 (1800)	6.3 (830)	2.4 (320)	1.3 (170)	1.1 (150)

*These values are approximate characteristic values based on MC-147.
**This slurry is 12.7% solids as in McGlammery, et al. (MC-147).

Assumptions:

a) 105% stoichiometric lime
b) 3.5% sulfur in coal (dry basis)
c) 12% Ash (as fired basis)
d) 28 MJ/kg (as fired basis)(12,000 Btu/lb)
e) 92% of sulfur in coal evolves as SO_2
f) 99% removal of particulates in ESP prior to absorption
g) 90% SO_2 removal
h) pond evaporation equals rainfall
i) 75% of ash evolves as fly ash
j) first stage S/L separator produces 40% solids sludge
k) ponding or second stage S/L produces 60% solids sludge
l) 10% solids recirculating slurry

to the effluent hold tank. If the waste solids are to be disposed of off-site, it is usually more economical to dewater them further by vacuum filtration and/or centrifugation to produce a final solids content of 60-70 weight percent. The filter cake is then transported to the disposal site. The filtrate and/or centrate are recycled to the effluent hold tank. When disposed of on-site, the clarifier underflow is pumped to large settling ponds where additional solids settling occurs. Clear supernatant is withdrawn from the pond and recycled to the scrubbing loop.

Water makeup is required to replace water losses due to evaporation in the absorber, and to occlusion in the solid waste. In certain cases, the water occluded in the solid waste can have an impact on surrounding water quality. This aspect of FGD-associated water impacts, however, is addressed in the report on solid waste impact (RO-359). Large influxes of rain water into the settling pond are handled by increasing the flow of pond supernatant back to the hold tank and decreasing the water makeup rate.

A purge stream may, however, be required from a lime FGD system because of several possible process problems. Fresh water must be added to the system for pump seals and demister washing. When the boiler is turned down so that flue gas flow is decreased, fewer solids are produced and, thus, the liquor bleed stream is reduced. This decreases the amount of water lost from the system. The amount of fresh water required for pump seals and demister washing does not decrease, however, so that a water imbalance necessitating a purge may occur. Changes in operation from design conditions, such as operating at lower sulfur coal or oxidizing sulfite sludge to sulfate (which can be dewatered more easily) could also decrease the amount of water lost and necessitate a purge. In addition, catastrophic occurrences could necessitate a quick blowdown to prevent scaling. Operator error could also result in a purge. In many cases, the need for a purge can be avoided by proper design and operation of the process. Treatment technologies are available to handle these purge streams.

5.2.3 Purge Characteristics

The possible purge from the lime system would have the same quality as the recirculated clarifier supernatant. This liquor differs from system to system due to coal type, fly ash collection facilities, and scrubber operation. The liquor is saturated with dissolved calcium sulfite and/or calcium sulfate salts. Sodium and chloride ions are also present in high concentrations. Most trace elements and toxic species are controlled below 1 mg/ℓ (BO-203).

5.3 Characterization of Effluents from the Limestone Wet Scrubbing Process

A description of the Limestone Wet Scrubbing Process is given in Section 3.3. The following section presents a closer examination of the water system. A water balance for one of four equivalent scrubbing trains for a base case 500 MW power plant is given. An examination of the water system flows permits the conclusion that there are no wastewater streams in the limestone wet scrubbing process in normal operation.

5.3.1 Base Case Water Balance

In calculating an example water balance to illustrate the water flows in a limestone wet scrubbing system, Radian chose a base case for ease of comparison. The base case is calculated for one of four equivalent scrubbing systems for a 500 MW power plant burning 3.5 percent sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb). The detailed calculations are presented in Appendix A. Figure 5.3-1 illustrates the water system. The results of the water balance calculations are presented in Table 5.3.1. Several of the circulating flow rates are characteristic flows rather than design flows, as indicated.

Flue gas first enters an ESP where 99 percent of the fly ash particulates are removed and sluiced to the ash pond. The flue gases, which contain some percentage water, then enter the absorber where SO_2 is removed. The gases are cooled and saturated by evaporation of water from the circulating limestone slurry. The saturated flue gas exits the absorber, is reheated for proper bouyancy and discharged to the stack. The absorber effluent is sent to a hold tank where calcium sulfite and sulfate crystals are precipitated. A 10-15 weight percent slurry is recirculated to the absorber. A bleed stream is sent to solids dewatering. The supernatant from the solid liquid separator is recycled to the hold tank. The underflow is typically 30-40 percent solids by weight. Either second stage solid-liquid separation or a settling pond is used to recover more of the liquor. This liquor is then recycled to the hold tank. The final sludge is typically 50-60 percent solids in a settling tank or 60-70 percent solids if a vacuum filter and/or centrifuge is used. Makeup water is therefore required to compensate for loss due to evaporation in the absorber and to occlusion in the solid waste.

5.3.2 Description of the Water System

The water systems in the limestone wet scrubbing process can be designed to operate in closed loops as exemplified by the limestone scrubbing system at Central Illinois Light Company's Duck Creek No. 1A and Kansas City Power and Light's La Cygne Station (PE-259). In a closed system, there are no aqueous

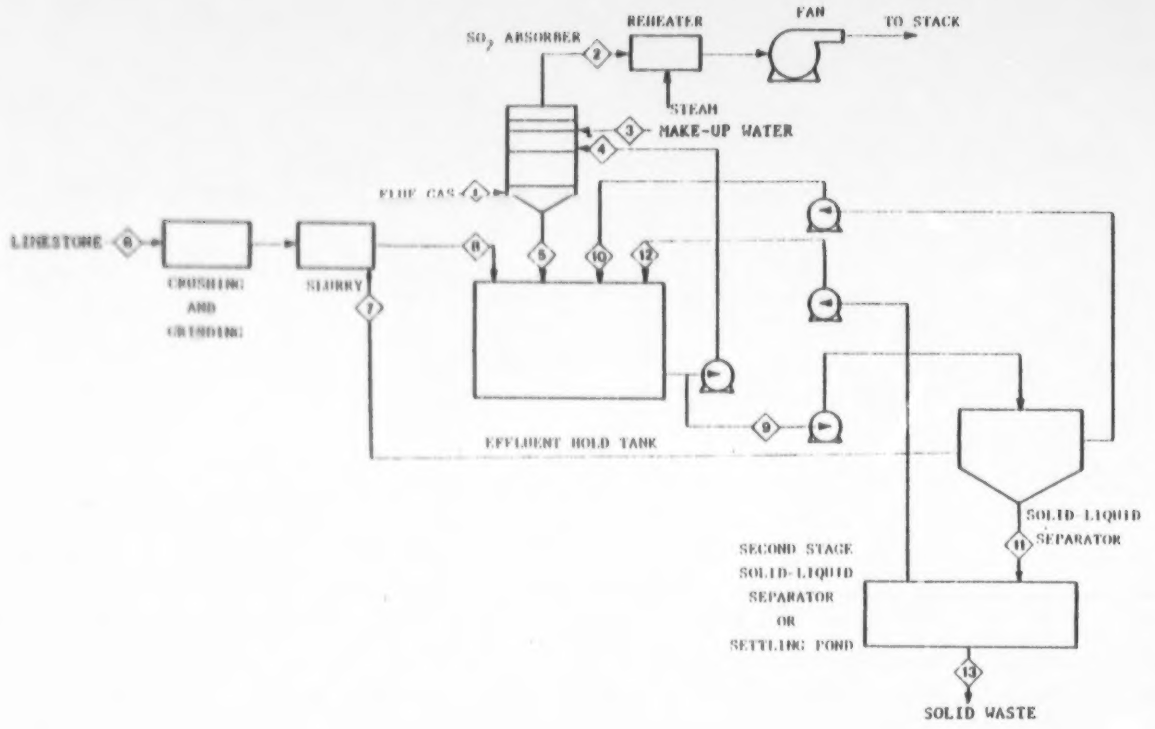

SO$_2$ ABSORBER
REHEATER
FAN
TO STACK
STEAM
MAKE-UP WATER
FLUE GAS
LIMESTONE
CRUSHING AND GRINDING
SLURRY
EFFLUENT HOLD TANK
SOLID LIQUID SEPARATOR
SECOND STAGE SOLID-LIQUID SEPARATOR OR SETTLING POND
SOLID WASTE
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Figure 5.3-1. Process flow diagram limestone wet scrubbing process.

TABLE 5.3-1. WATER BALANCE: LIMESTONE WET SCRUBBING PROCESS

(one of four equivalent scrubbing trains)

Stream No.	1	2	3	4*	5*	6	7**	8	9	10	11	12	13
Description	Gas to Absorber	Gas to Air Heater	Make-up Water to Heater	Recycle Slurry to Absorber	Slurry to Hold Tank	Limestone Feed	First Stage S/L Overflow to Slurry Feed Tank	Limestone Slurry to Hold Tank	Slurry Bleed Stream	Solid-Liquid Separator Overflow to Hold Tank	Solid-Liquid Separator Underflow to Pond or Filtration	Second Stage S/L Separator Overflow to Hold Tank	Solid Waste
Stream Rate, kg/s (lb/min)	160 (21,000)	170 (22,000)	9.1 (1200)	1500 (200,000)	1500 (200,000)	1.3 (170)	0.76 (100)	2.0 (270)	17 (2200)	11 (1500)	4.3 (570)	1.4 (180)	2.9 (390)
Water, kg/s (lb/min)	7.6 (1000)	15 (200)	9.1 (1200)	1400 (180,000)	1400 (180,000)	--	0.76 (100)	0.76 (100)	15 (2000)	12 (1600)	2.6 (350)	0.64 (85)	1.3 (170)

*These values are approximate characteristic values.

**This stream is 6.3% solids as in McGlamerry, et al. (MC-147).

Assumptions:

a) 120% stoichiometric limestone
b) 3.5% sulfur in coal (dry basis)
c) 12% ash (as fired basis)
d) 28 MJ/kg (as fired basis)(12,000 Btu/lb)
e) 12% of sulfur in coal evolves as SO_2
f) 99% removal of particulates in ESP prior to absorption
g) 90% SO_2 removal
h) pond evaporation equals rainfall
i) 75% of ash evolves as fly ash
j) first stage S/L separator produces 40% solids sludge
k) ponding or second stage S/L produces 60% solids sludge
l) 10% solids recirculating slurry

discharges. The limestone slurry is recirculated between the effluent hold tank and the absorber. A bleed stream is taken for solids dewatering and disposal. The solids are generally removed in a clarifier underflow as a 30-40 weight percent sludge. The clarifier supernatant is recycled to the effluent hold tank. If the waste solids are to be disposed of on-site, the sludge is usually ponded and allowed to settle. A final sludge of 50-60 weight percent solids can be achieved with the clear supernatant liquor from the sludge pond recycled to the effluent hold tank. If the waste solids are to be disposed of off-site, it is usually more economical to dewater them further by vacuum filtration and/or centrifugation to produce a final solids content of 60-70 weight percent. The filter cake is then transported to the disposal site. The filtrate and/or centrate are recycled to the effluent hold tank.

A net water makeup is required to replace water losses in the process due to evaporation in the absorber, and to occlusion in the solid waste. In areas where rainfall exceeds solar evaporation rates, more supernatant liquid is returned from the settling pond than sent to the pond. This influx of water into the closed loop limestone scrubbing system can be offset by reducing the raw water makeup to the process. In certain cases, the water occluded in the settled solid wastes can have an impact on surrounding water quality. This aspect of the FGD associated water impacts, however, is addressed in the report on solid waste impact (RO-359).

A purge stream may be required from a limestone FGD system because of several possible process problems. Fresh water must be added to the system for pump seals and demister washing. When the boiler is turned down so that flue gas flow is decreased, fewer solids are produced and, thus, the liquor bleed stream is reduced. This decreases the amount of water lost from the system. The amount of fresh water required for pump seals and demister washing does not decrease, however, so that a water imbalance necessitating a purge may occur. Changes in operation from design conditions, such as operating at lower sulfur coal or oxidizing sulfite sludge to sulfate (which can be dewatered more easily) could also decrease the amount of water lost and necessitate a purge. In addition, catastrophic occurrences could necessitate a quick blowdown to prevent scaling (BO-203). Operator error could also result in a purge. In many cases, the need for a purge can be avoided by proper design and operation of the process. Treatment technologies are available to handle these purge streams.

5.3.3 Purge Characteristics

The possible purge from the limestone system would have the same quality as the recirculated clarifier supernatant. This liquor differs from system to system due to coal type, fly ash collection facilities, and scrubber operation. The liquor is saturated with dissolved calcium sulfite and/or calcium sulfate salts. Sodium and chloride ions are also present in high concentrations. Most trace elements and toxic species are controlled below 1 mg/ℓ (BO-203).

5.4 Characterization of Wastewaters From the Wellman-Lord Sulfite Scrubbing Process

A description of the Wellman-Lord process is givn in Section 3.4. The following section presents a closer examnation of the water system. A water balance is given for a base ase condition for one of four equivalent scrubbing trains appled to a 500 MW power plant. An examination of the water system hows that one wastewater stream is associated with the prescrubbing of flue gases. This stream contains particulates and chloride, but is insignificant in comparison to the wastewater streams fom power plants.

5.4.1 Base Case Water Balance

In calculating an example water balance to illustrate the water flows in a Wellman-Lord sulfite scrubbing system, Radian has chosen a base case for ease of comparison. The case is calculated for one of four equivalent scrubbing systems for a 500 MW power plant burning 3.5 percent sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb). The detailed calculations are presented in Appendix A. Figure 5.4-1 illustrates the water system. The results of the water balance calculations are presented in Table 5.4-1. Several of the circulating flow rates are characteristic flows rather than design flows, as indicated.

Flue gas first enters an ESP where 99 percent of the particulates are removed and sluiced to the ash pond. The hot flue gases then enter a prescrubber, where the remaining particulates and 99 percent of the chlorides are scrubbed from the gases. The gases are cooled and saturated by the evaporation of water in the prescrubber. A bleedstream is taken from the scrubber circulating slurry to maintain desired suspended and dissolved solids levels. This stream is then routed to the ash pond. The scrubbed flue gases enter the absorber where SO_2 is removed and absorbed into a circulating sodium sulfite solution. The cleaned gases exit the absorber, are reheated, and exhausted to the stack. No additional evaporation of water takes place in the absorber. The absorber effluent is sent to double effect evaporators where SO_2 and water are driven off. The overhead is passed through condensers and most of the water is condensed. The water is then sent to the dissolving tank for recombination with the evaporator underflow and make-up sodium carbonate. The resulting solution is used for absorber feed. The product SO_2 stream contains 5-10 percent water.

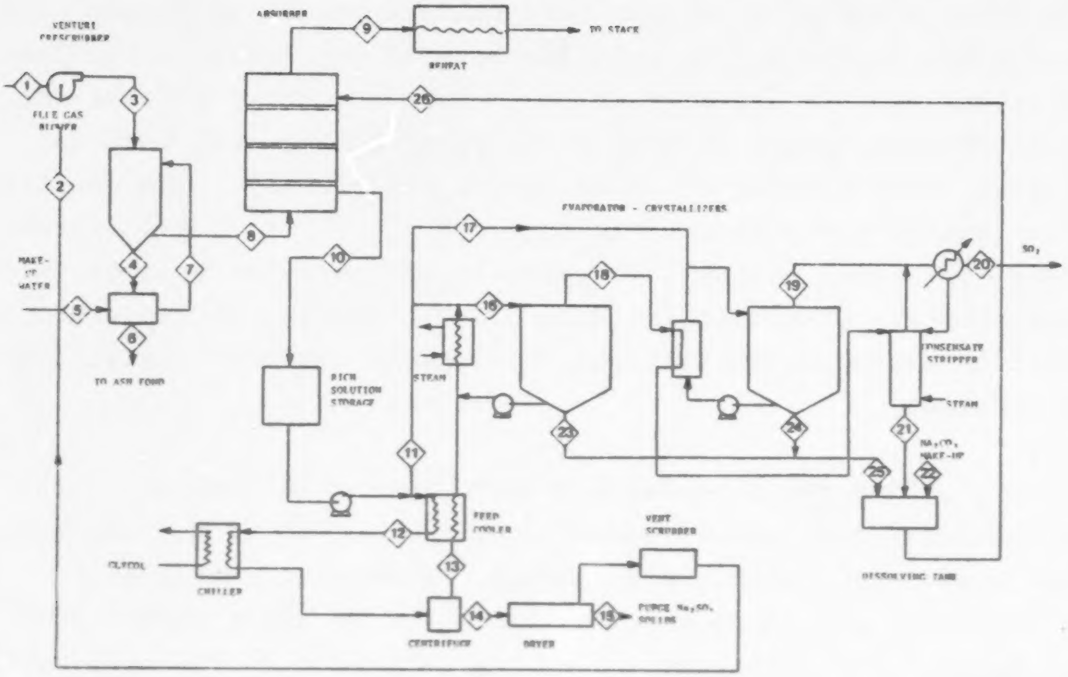
VENTURI PRESCRUBBER
ABSORBER
9
REHEAT
TO STACK
1
FLUE GAS BLOWER
3
26
2
8
MAKE-UP WATER
4
7
5
6
TO ASH POND
10
EVAPORATOR - CRYSTALLIZERS
17
18
19
20
SO_2
16
STEAM
RICH SOLUTION STORAGE
CONDENSATE STRIPPER
STEAM
23
24
21
Na_2CO_3 MAKE-UP
11
25
22
FEED COOLER
12
GLYCOL
CHILLER
13
VENT SCRUBBER
DISSOLVING TANK
14
15
PURGE Na_2SO_4 SOLIDS
CENTRIFUGE
DRYER

Figure 5.4-1. Process flow diagram for the Wellman-Lord sulfite scrubbing process.

TABLE 5.4-1. WATER BALANCE: WELLMAN-LORD SULFITE SCRUBBING PROCESS

(one of four equivalent scrubbing trains)

Stream No. Figure 5.4-1	1	2	3	4*	5	6	7*	8	9	10*	11	12	13
Description	Flue Gas to Scrubber	Dryer Gas to Scrubber	Total Gas to Scrubber	Scrubber Liquor Product	Scrubber Make-up Water	Slurry to Ash Pond	Circulating Slurry to Scrubber	Gas to Absorber	Gas to Reheat	Absorber Liquor Product	Absorber Liquor to Evaporator	Absorber Liquor to Evaporator	Centrate to Evaporator
Stream Rate, kg/s (lb/min)	160 (21,000)	5.9 (780)	170 (22,000)	300 (40,000)	9.1 (1200)	0.91 (120)	300 (40,000)	170 (23,000)	170 (23,000)	11 (1500)	9.8 (1300)	1.7 (230)	1.5 (230)
Water, kg/s (lb/min)	7.6 (1000)	0.083 (11)	7.6 (1000)	290 (38,000)	9.1 (1200)	0.83 (110)	290 (38,000)	16 (2100)	16 (2100)				

Stream No. Figure 5.4-1	14	15	16	17	18	19	20	21	22	23	24	25	26
Description	Purge Solids to Dryer	Purge ,Solids	First Effect Evaporator Feed	Second Effect Evaporator Feed	First Effect Evaporator Overhead	Second Effect Evaporator Overhead	SO_2 Product Stream	Condensate	Make-up Na_2CO_3	First Effect Slurry	Second Effect Slurry	Slurry to Dissolving Tank	Feed Solution to Absorber
Stream Rate, kg/s (lb/min)	0.17 (23)	0.083 (11)	6.7 (890)	4.5 (590)	4.0 (530)	2.6 (350)	0.71 (95)	6.2 (820)	0.22 (30)	2.6 (350)	1.8 (240)	4.5 (590)	11 (1500)
Water kg/s (lb/min)	0.083 (11)						0.071 (9.5)	6.2 (820)	0.16 (21)				

*Characteristic values.

Assumptions:

a) 3.5% sulfur in coal (dry basis)
b) 12% ash (as fired basis)
c) 28 MJ/kg (as fired basis)(12,000 Btu/lb)
d) 92% of sulfur in coal evolves as SO_2
e) 99% removal of particulates in ESP prior to scrubbing
f) 90% SO_2 removal
g) 75% of ash evolves as fly ash
h) 15% of circulating slurry is sent to purge treatment
i) double effect evaporator
j) 60% overhead in each effect
k) 10% H_2O in product SO_2 stream

Approximately 15 percent of the absorber effluent is sent to purge treatment to remove sulfates and other impurities from the system. The purge stream is cooled and centrifuged to remove 70 percent sodium sulfate and 30 percent sodium sulfite crystals. The centrate is routed to the evaporators and the crystals are dried. The dryer gases are routed to the prescrubber to evaporate additional water therein. The dryer gases and evaporated water exit through the absorber and stack with the flue gases.

Water losses in the Wellman-Lord process are due to evaporation in the prescrubber, prescrubber blowdown, loss in the product SO_2 stream, and drying of purge solids. In addition, the blowdown of the condenser cooling water system has evaporative and drift losses.

5.4.2 Description of Water System

The water systems in the Wellman-Lord sulfite scrubbing process operate in closed loops. The prescrubbing system recirculates a slurry of approximately 5 weight percent solids. This system requires a small blowdown to maintain the suspended and dissolved solids at desired levels. The absorber also operates in a closed loop, with 85 percent of the scrubbing solution being circulated through double effect evaporators. Regeneration of the scrubbing solution and production of a concentrated SO_2 stream are thus accomplished. The remaining 15 percent is sent to purge treatment. The liquor is then either recycled to the evaporator loop or is evaporated into the dryer combustion gases which exit through the scrubber and absorber. Thus, the only effluents from the Wellman-Lord process are the prescrubber blowdown and the solid by-product. The solid is currently sold to the paper industry. The prescrubber blowdown is the only wastewater.

5.4.3 Blowdown Characteristics

The size of the prescrubber blowdown for the base case calculation was limited by the particulate loading (Appendix A). Because 99 percent of the particulates are removed in an ESP prior to the gases' entering the scrubber, the size of this stream is relatively small compared to the current power plant ash sluicing requirement. The quality of this stream will be similar to that of ash sluicing waters (see Table 5.1-1) with the addition of high chloride concentrations (about 10,000 mg/ℓ). The effect on receiving streams and treatment technologies will be discussed in Section 6.

5.5 Characterization of Wastewaters From the Magnesia Slurry Absorption Process

A description of the magnesia slurry absorption process is given in Section 3.5. The following section presents a closer examination of the water system. A water balance is given for a base case condition for one of four equivalent scrubbing trains applied to a 500 MW power plant. An examination of the water system indicates that the wastewater streams associated with the magnesia slurry process are insignificant in comparison to current power plant wastewaters. The wastewater streams include: 1) a prescrubber blowdown for suspended and dissolved solids control, and 2) an intermittent purge from the absorber loop to remove impurities which enter the system in makeup streams and as products of oxidation.

5.5.1 Base Case Water Balance

In calculating an example water balance to illustrate the water flows in the magnesia slurry absorption process, Radian has chosen a base case for ease of comparison. The base case is calculated for one of four equivalent scrubbing systems for a 500 MW power plant burning 3.5 percent sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb). The detailed calculations are presented in Appendix A. Figure 5.5-1 illustrates the water system. The results of the water balance calculations are presented in Table 5.5-1. Several of the circulating flow rates are characteristic flows rather than design flows, as indicated.

In the magnesia slurry absorption process, flue gas first enters an ESP where 99 percent of the particulates are removed and sluiced to the ash pond. The hot flue gases then enter a prescrubber where the remaining particulates and 99 percent of the chlorides are scrubbed from the gases with water. The gases are also cooled and saturated by the evaporation of water from the scrubbing slurry. A blowdown is taken from the circulating scrubbing solution to maintain the desired suspended and dissolved solids levels. This stream is then routed to the ash pond. The scrubbed flue gases enter the absorber, where SO_2 is removed and absorbed into a circulating magnesia slurry. The cleaned gases exit the absorber, are reheated, and exhausted to the stack. No additional evaporation of water takes place in the absorber. The absorber effluent is recycled. A bleed stream is sent to regeneration; makeup slurry is added to the recirculated slurry prior to entering the absorber. The bleed stream is passed through thickeners to produce a 40 weight percent solids slurry. It is then thermally treated to release some water of hydration, and centrifuged. The liquor from each of these operations is recycled to the slurry tank. The centrifuge cake is then dried with the dryer gases being exhausted to the stack. The magnesia sulfate crystals are calcined to release SO_2 and regenerate MgO. The regenerated MgO is recycled to the slurry tank, while the SO_2 product gas is sent to SO_2 conversion.

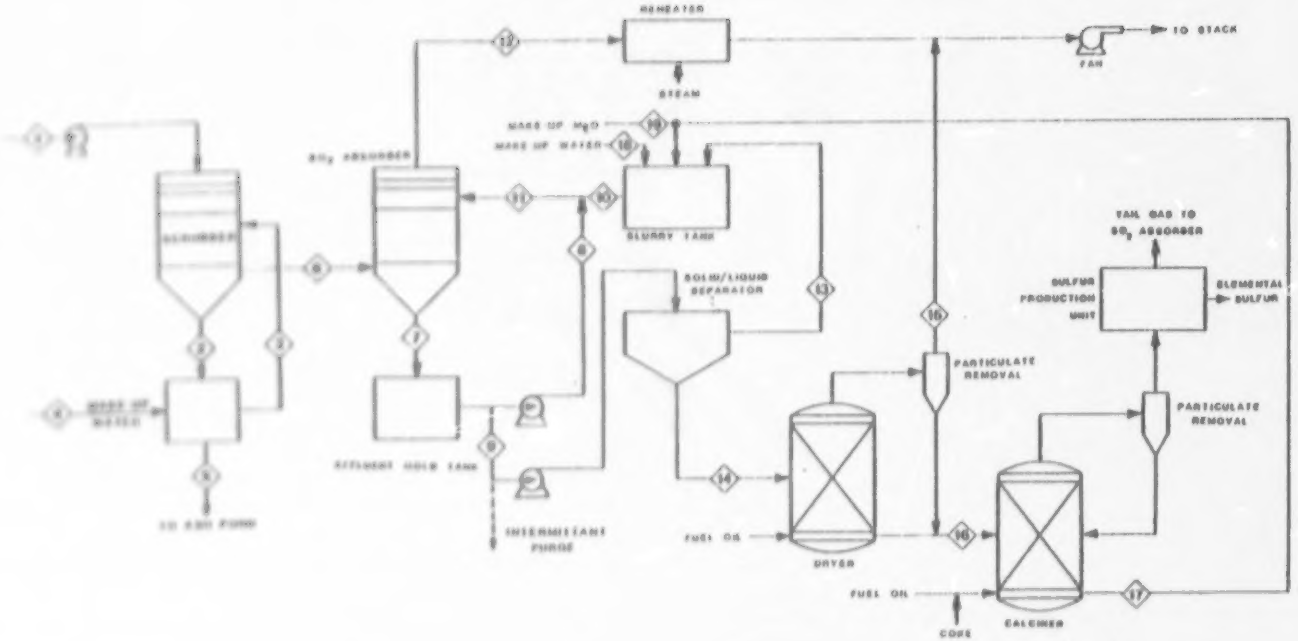
REHEATER
STEAM
FAN
TO STACK
SCRUBBER
MAKE UP WATER
TO ASH POND
SO2 ABSORBER
EFFLUENT HOLD TANK
INTERMITTENT PURGE
MAKE UP MgO
MAKE UP WATER
SLURRY TANK
SOLID/LIQUID SEPARATOR
PARTICULATE REMOVAL
DRYER
FUEL OIL
FUEL OIL
COKE
CALCINER
PARTICULATE REMOVAL
SULFUR PRODUCTION UNIT
TAIL GAS TO SO2 ABSORBER
ELEMENTAL SULFUR
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Figure 5.5-1. Process flow diagram for the magnesia slurry absorption process.

TABLE 5.5-1. WATER BALANCE: MAGNESIA SLURRY ABSORPTION PROCESS
(one of four equivalent scrubbing trains)

Stream No. Figure 5.5-1	1	2*	3*	4	5	6	7*	8*	9	10*	11*	12**	13
Description	Gas to Scrubber	Scrubber Slurry to Surge Tank	Scrubber Recycle Slurry	Make-up Water to Scrubber	Scrubber Effluent to Ash Pond	Gas to SO_2 Absorber	Absorber Effluent to Hold Tank	Recycle Slurry	Slurry to S/L Separator	Make-up MgO Slurry	Recycle Slurry to SO_2 Absorber	Gas to Air Heater	S/L Overflow to Slurry Tank
Stream Rate, kg/s (lb/min)	160 (21,000)	300 (40,000)	300 (40,000)	8.3 (1100)	0.91 (120)	170 (22,000)	440 (58,000)	410 (54,000)	34 (4500)	32 (4300)	430 (57,000)	170 (22,000)	32 (4300)
Water, kg/s (lb/min)	7.6 (1000)	290 (38,000)	290 (38,000)	8.3 (1100)	0.83 (110)	15 (2000)	390 (52,000)	360 (48,000)	31 (4100)	29 (3900)	390 (52,000)	15 (2000)	32 (4300)

Stream No. Figure 5.5-1	14	15*	16	17	18	19
Description	S/L Underflow to Dryer	Dryer Gas to Stack	Feed to Calciner	Recycle MgO	Make-up Water to Slurry Tank	Make-up MgO
Stream Rate, kg/s (lb/min)	2.1 (280)	17 (2300)	1.4 (180)	0.53 (70)	0.71 (94)	0.037 (5)
Water, kg/s (lb/min)	0.12 (15)	0.71 (94)	--		0.71 (94)	

*These values are approximate characteristic stream values.
**This value is calculated based on gas flow rates in MC-147.

Assumptions:

a) 105% stoichiometric magnesia
b) 3.5% sulfur in coal (dry basis)
c) 12% ash (as fired basis)
d) 28 MJ/kg (as fired basis)(12,000 Btu/lb)
e) 75% of ash evolves as fly ash
f) 92% of sulfur in coal evolves as SO_2
g) 99% removal of particulates in ESP prior to scrubbing
h) 90% SO_2 removal
i) pond evaporation equals rainfall
j) first stage S/L separator produces a 40% solids slurry
k) centrifuge produces a 95% solids cake
l) 10% solids recirculating slurry

Water losses in the magnesium slurry process are due to evaporation in the prescrubber, prescrubber blowdown, and drying of solids in the regeneration loop.

5.5.2 Description of the Water System

The water systems in the magnesia slurry absorption process operate in closed loops (MC-076). The prescrubbing system recirculates a slurry of approximately 5 weight percent solids. This system requires a small blowdown to maintain the suspended and dissolved solids at desired levels. Most of the absorber effluent is recirculated; the regenerative bleedstream operates in closed loops. The liquor from the thickeners and centrifuge are recycled to the slurry tank. Water lost in drying the crystals exits as a vapor through the stack.

Thus, the only continuous wastewater associated with the magnesia slurry process is the prescrubber blowdown.

A second possible wastewater is an intermittent purge from the absorber loop. Impurities enter the system in makeup water and makeup MgO, and are concentrated. In addition, sulfite ion is oxidized to sulfate. Because of developmental status of this process, little information is available to predict the effect of long-term build-up of $MsSO_4$. McGlammey, et al (MC-076) have estimated that a purge of approximately 63 cm^3/s (1 gal/min) for the base case might be necessary. These authors have suggested several possible treatment technologies which will be discussed in detail in Section 6.

5.5.3 Blowdown Characteristics

The size of the prescrubber blowdown for the base case calculation was limited by the particulate loading (Appendix A). Because 99 percent of the particulates are removed in an ESP prior to the gases entering the scrubber, the size of this stream is relatively small compared to the current power plant ash sluicing requirement. The quality of this stream will be similar to that of ash sluicing waters (see Table 5.1-1) with the addition of high chloride concentrations (about 10,000 mg/ℓ). The effect on receiving streams and treatment technologies will be discussed in Section 6.

5.5.4 Purge Characteristics

The impurities in the purge will come from makeup water and makeup MgO. The MgO can be expected to contain silica, ferric oxide, alumina, chloride, sulfate, and calcium oxide impurities. Makeup water quality may vary considerably, but may contain calcium oxide, sulfate, chloride and trace metal impurities. McGlammery, et al (MC-076) assume that the purge will be a clarified solution of approximately 1.2 percent $MgSO_3$ and 15 percent $MgSO_4$.

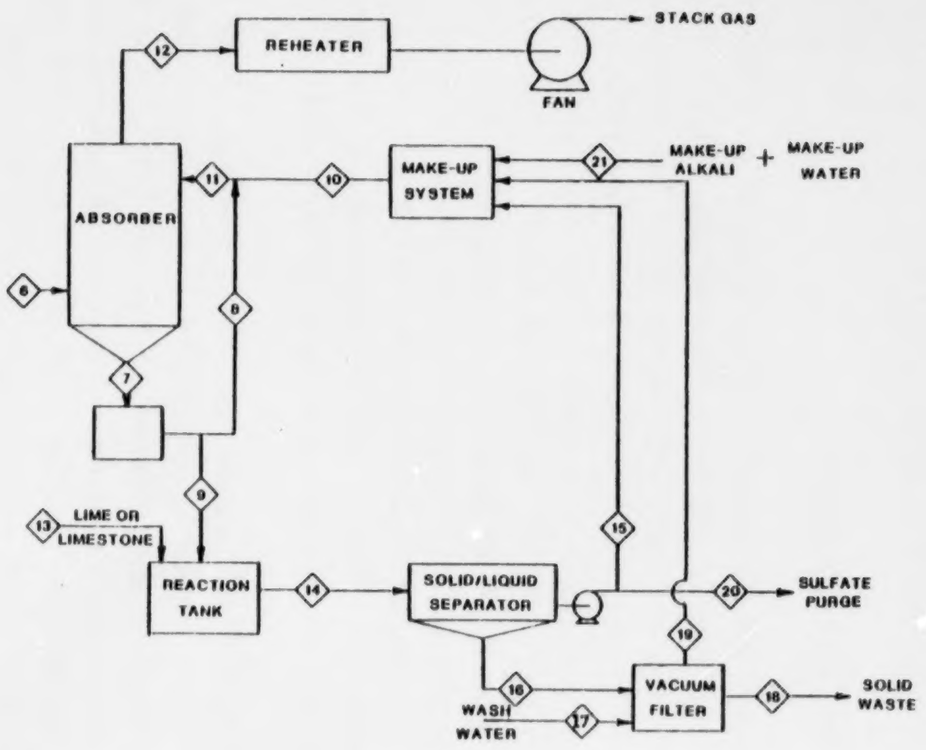
5.6 Characterization of Wastewaters from the Double Alkali Wet Scrubbing Process

A description of the double alkali wet scrubbing process is given in Section 3.6. The following section presents a closer examination of the water system. A water balance is given for a base case condition for one of four equivalent scrubbing trains applied to a 500 MW power plant. An examination of the water system shows that there are no continuous wastewater streams associated with the process. If a prescrubber is used for chloride removal, there will be a wastewater stream consisting of prescrubber blowdown. This stream would contain particulates and chlorides, but is insignificant in comparison to current wastewater streams from power plants.

5.6.1 Base Case Water Balance

In calculating an example water balance to illustrate the water flows in a double alkali wet scrubbing system, Radian has chosen a base case for ease of comparison. The base case is calculated for one of five equivalent scrubbing systems for a 500 MW power plant burning 3.5 percent sulfur coal with an average heating value of 28 MJ/kg (12,000 Btu/lb). The detailed calculations are presented in Appendix A. Figure 5.6-1 illustrates the water system. The results of the water balance calculations are presented in Tables 5.6-1 and 5.6-2 for lime and limestone regenerant, respectively. Several of the circulating flow rates are characteristic flows and not design flows.

In the double alkali wet scrubbing process, flue gases containing water first enter an ESP where 99 percent of the particulates are removed and sluiced to the ash pond. The hot flue gases then enter the absorber, where SO_2 is removed and absorbed into a circulatory sodium sulfite solution. The gases are cooled and saturated by evaporation of water from the scrubbing slurry. This saturation may occur in a presaturation section of the absorber. The cleaned gases exit the absorber, are reheated and exhausted to the stack. A bleed stream is taken off the absorber effluent and sent to a reaction tank. There either lime or limestone regenerant is added to form the


STACK GAS
REHEATER
FAN
12
MAKE-UP SYSTEM
21
MAKE-UP ALKALI
+
MAKE-UP WATER
ABSORBER
11
10
6
8
7
9
13
LIME OR LIMESTONE
REACTION TANK
14
SOLID/LIQUID SEPARATOR
15
20
SULFATE PURGE
19
16
VACUUM FILTER
18
SOLID WASTE
WASH WATER
17

Note: Streams ① → ⑤ not shown.

Figure 5.6-1. Process flow diagram for double-alkali wet scrubbing.

TABLE 5.6-1. WATER BALANCE: DOUBLE ALKALI WET SCRUBBING, LIME REGENERANT

(One of four equivalent scrubbing trains)

Stream No. Figure 5.6-1	6	7	8	9	10	11	12	13
Description	Gas to Absorber	Absorber Liquor Product	Recycle Liquor	Bleed Stream to Reaction Tank	Make-up Liquor	Feed Liquor to Absorber	Gas to Reheat	Lime
Stream Rate, kg/s (lb/min)	160 (21,000)	250 (33,000)	220 (29,000)	30 (3900)	36 (4800)	250 (33,000)	170 (22,000)	21 (280)
Water, kg/s (lb/min)	7.6 (1000)						1.5 (200)	

Stream No. Figure 5.6-1	14	15	16	17	18	19	20	21*
Description	Slurry to S/L Separator	S/L Overflow to Make-up System	S/L Underflow to Vacuum Filtration	Filter Cake Wash Water	Solid Waste	Centrate to Make-up System	Sulfate Purge	Make-up Water and Alkali
Stream Rate, kg/s (lb/min)	32 (4200)	30 (3700)	3.9 (520)	2.3 (300)	2.6 (350)	3.6 (470)		4.9 (650)
Water, kg/s (lb/min)	30 (4000)		2.4 (320)	2.3 (300)	1.1 (150)			4.9 (650)

* 0.027 kg/s (3.6 lb/min) alkali

Assumptions:

a) 105% lime stoichiometry
b) 3.5% sulfur in coal (dry basis)
c) 12% ash (as fired basis)
d) 28 MJ/kg (as fired basis)(12,000 Btu/lb)
e) 92% of sulfur in coal evolves as SO_2
f) 99% removal of particulate in ESP prior to scrubbing
g) 90% SO_2 removal
h) pond evaporation equals rainfall
i) 75% of ash evolves as fly ash
j) first stage S/L separator produces 40% solids underflow
k) second stage S/L or ponding produces 60% solids sludge
l) solid waste washed with 2 displacement washes

TABLE 5.6-2. WATER BALANCE: DOUBLE ALKALI WET SCRUBBING, LIMESTONE REGENERANT

(One of four equivalent scrubbing trains)

Stream No.	6	7	8	9	10	11	12	13
Description	Gas to Absorber	Absorber Liquor Product	Recycle Liquor	Bleed Stream to Reaction Tank	Make-up Liquor	Feed Liquor to Absorber	Gas to Reheat	Limestone
Stream Rate, kg/s (lb/min)	160 (21,000)	250 (33,000)	210 (28,000)	32 (4300)	39 (5200)	250 (33,000)	170 (22,000)	4.3 (570)
Water, kg/s (lb/min)	7.6 (1000)						15 (2000)	3.0 (400)

Stream No.	14	15	16	17	18	19	20	21*
Description	Slurry to S/L Separator	S/L Overflow to Make-up System	S/L Underflow to Vacuum Filtration	Filter Cake Wash Water	Solid Waste	Centrate to Make-up System	Sulfate Purge	Make-up Alkali
Stream Rate, kg/s (lb/min)	36 (4800)	33 (4300)	4.2 (550)	2.5 (330)	2.9 (390)	3.8 (500)		3.3 (430)
Water, kg (lb/min)	35 (4600)		2.5 (330)	2.5 (330)	1.3 (170)			3.3 (430)

* 0.027 kg/s (3.6 bl/min) alkali

Assumptions:

a) 120% limestone stoichiometry
b) 3.5% sulfur in coal (dry basis)
c) 12% ash (as fired basis)
d) 28 MJ/kg (as fired basis)(12,000 Btu/lb)
e) 92% of sulfur in coal evolves as SO_2
f) 99% removal of particulate in ESP prior to scrubbing
g) 90% SO_2 removal
h) pond evaporation equals rainfall
i) 75% of ash evolves as fly ash
j) first stage S/L separator produces 40% solids underflow
k) second stage S/L or ponding produces 60% solids sludge
l) solid waste washed with 2 displacement washes

solid waste and to regenerate the scrubbing solution. The reaction tank slurry is sent to a solid/liquid separator, where the supernatant is recycled to the makeup system. The 40 percent solids underflow is sent to a vacuum filter. The filter cake is washed to recover soluble sodium. The filtrate and wash water are recycled to the makeup system. The final filter cake is 60 percent solids. The main portion of the absorber effluent is combined with regenerated liquor and recirculated as absorber feed.

Thus, water losses in the double alkali wet scrubbing process are due to evaporation in the absorber and occlusion in the solid waste. If a prescrubber is used, prescrubber blowdown would be an additional water loss.

5.6.2 Description of the Water Systems

The water systems in the double alkali wet scrubbing process operate in closed loops. If a prescrubber is required, this system would require a small blowdown to maintain the suspended and dissolved solids at desired levels. Approximately 90 percent of the absorber effluent is recirculated. A bleed stream is regenerated with lime or limestone in a reaction tank, but all supernatant liquors from the solid liquid separations are recycled to the makeup system. The water occluded in the solid waste can have an impact on surrounding water quality in some cases, but this aspect of the FGD-associated water impacts is being addressed in the report on solid waste impact (RO-359).

Thus, there are no continuous wastewaters associated with the double alkali system in normal operation. Scrubber liquor would have to be treated for sulfate removal, as discussed in Section 6.2. In some cases, a purge may be necessary to control the concentration of nonsulfur/calcium species in the system. Such a purge should not be required in normal operation, however. As discussed earlier, a prescrubber may be required under certain conditions, and would necessitate a blowdown stream.

5.6.3 Purge Characteristics

The sulfate purge characteristics are the same as those of the recirculated clarifier overflow. The purge contains sodium sulfate, sodium sulfite, and soluble nonsulfur/calcium species in varying amounts. The nonsulfur/calcium species enter the system in the lime and/or limestone, and in the makeup water. In systems having common prescrubber and absorber circulating loops, these species enter with the fly ash and flue gas. The soluble nonsulfur/calcium contaminant in highest concentration is probably sodium chloride. Contamination results from the absorption of HCl from the flue gas.

5.6.4 Blowdown Characteristics

If a prescrubber is required for chloride removal, the size and quality of the blowdown stream would be comparable to that of blowdowns from the Wellman-Lord and Magnesia Slurry systems.

5.7 Characterization of Wastewaters from the Physical Coal Cleaning Process

The physical coal cleaning plants vary widely in types of processes used and in plant layout. Generally, water systems in new plants operate in closed loops (LE-218). Solid refuse is allowed to settle in ponds and the pond water is recycled for reuse in the processing areas. Some water is occluded in the solid refuse, while some is lost due to thermal drying operations or increased coal moisture content.

Coal cleaning by the wet process method uses about .0063 to .0083 m^3 of water per kg of coal (1,500 to 2,000 gallons of water per ton of coal processed). Contaminants consist of suspended solids, which are chiefly fine clay and coal, and dissolved solids, which may contain iron, aluminum, calcium, magnesium, sodium, and potassium. Water effluents may also contain surface-active organic compounds such as alcohols or kerosene, which are added in some coal cleaning plants to enhance frothability in the process. In a modern plant, all liquid waste streams are routed to holding ponds to allow settling of the suspended solids. The clear supernatant liquid is then recycled to the process. Thus, no liquid effluents result directly from the cleaning process (LE-218). Water runoff from refuse piles, however, may contribute to water pollution. Water contaminants in refuse pile runoff include sulfuric acid, sulfates, manganese, and iron in varying concentrations. For example, in effluents from four different refuse sites, the concentration of sulfates varied from 690-9500 mg/ℓ; of manganese, 3.5 to 120 mg/ℓ; and of iron, 6.2 to 3400 mg/ℓ (MA-411).

5.8 Characterization of Wastewaters from SO_2 Conversion Processes

The Allied Chemical Process for elemental sulfur production has insignificant water systems which operate in closed loops. However, the production of sulfuric acid requires significant quantities of cooling water.

Elemental Sulfur Production

The water system in the Allied Chemical Process includes steam generation in the sulfur condenser, and cooling water for compressor seals. Both of these systems have relatively small requirements (as discussed in Section 4). The blowdown from these systems is expected to be negligible in quantity. The quality will be similar to that of cooling tower blowdown streams or steam system blowdowns. The Allied Process, as presented by Allied (HU-051), appears to have no liquid effluents when operated on an SO_2 stream from an FGD process.

Sulfuric Acid Production

Product acid coolers require significant quantities of cooling water. The effluent stream would be a blowdown similar to those described in regard to power plant cooling tower blowdowns. However, the quantities required are one to two orders of magnitude less than those required by the power plant. The recirculating water flow for sulfuric acid cooling is about 0.32 m^3/s (500 gpm), compared to about 13 m^3/s (210,000 gpm) for the condenser cooling system of the base case 500 MW power plant. The blowdown rate typically ranges between 0.5 and 3.0% of the recirculating water flow (DO-051).

6.0 EXAMINATION OF PURGE CHARACTERISTICS AND APPLICABLE TREATMENT TECHNOLOGY

This section examines the FGD-associated wastewater streams in detail. This examination includes a characterization of the effluent composition, an assessment of effect on receiving stream water quality, and possible treatment technologies. The specific wastewater streams addressed in this section include:

- Magnesia Slurry Absorption Process Purge,
- Double Alkali Sulfate Purge,
- Prescrubber Blowdown,
- Cooling Water System Blowdown,
- Possible Lime/Limestone Purge, and
- Lime/Limestone/Double Alkali Solid Waste.

6.1 Magnesia Slurry Absorption Process Purge

The magnesia slurry scrubbing system operates in a closed loop, and is subject to buildup of impurities that could lead to corrosion and scaling problems. Impurities enter the system in makeup water and makeup MgO. In addition, some $MgSO_3$ is oxidized to $MgSO_4$. Because of the developmental stage of the process there is little information available to predict the effect of buildup of $MgSO_4$ in continuously operated, closed loop cycles. Some sulfate removal may prove necessary.

6.1.1 Purge Characteristics

The impurities in the purge will come from makeup water and makeup MgO. The MgO can be expected to contain silica, ferric oxide, alumina, chloride, sulfate, and calcium oxide impurities. Makeup water quality may vary considerably, but may contain calcium oxide, sulfate, chloride and trace metal impurities. McGlammery, et al (MC-076) assume that the purge will be a clarified solution of approximately 1.2 percent $MgSO_3$ and 15 percent $MgSO_4$.

6.1.2 Effect on Receiving Streams

This purge would not be discharged directly to a receiving stream. Economics make recovery of $MgSO_4$ desirable. Furthermore, the Federal Water Pollution Control Act Amendment of 1972 and the discharge limitations based on best practical control technology available in 1977 and best available technology economically achievable in 1983 would prohibit direct discharge.

6.1.3 Treatment Technology

McGlammery et al (MC-076) suggested three purge treatment techniques:

1) sending a sidestream to a deadend pond;

2) concentration of the mother liquor until $MgSO_4$ precipitates before sending to a deadend pond, or treating;

3) "dissolving $MgSO_3 \cdot 6H_2O$ slurry with a minimum amount of sulfur dioxide, filtering the insoluble impurities, then reprecipitating sulfite with makeup MgO. The resultant crystals would be filtered and returned to the system; the mother liquor would be evaporated to

> recover $MgSO_4$ and the supernatant of soluble impurities discarded."

The first technique would be the simplest and probably least expensive purge treatment. Evaporation ponds have been used in water management throughout the chemical and electric utility industries. Solar energy is used to evaporate water from the pond. Dissolved solids in the waste stream fed to the pond are, thus, concentrated and precipitated. The precipitated dissolved solids are usually allowed to accumulate at the bottom of the pond, but in some instances are periodically dredged from the pond. The pond may have to be lined to prevent seepage of dissolved chemicals into underground water supplies. If so lined, there will be no effluents from the pond. Applicability of evaporation ponds depends on the net evaporation rate (the gross evaporation rate minus rainfall). Geographical areas with less than 50 cm (20 in.) net evaporation rate are not suitable for ponds. Water in a purge stream treated in an evaporation pond is lost by evaporation; and is thus not available for reuse in the system. Magnesium contained in the purge is also lost from the system (EN-392, EN-394, MC-076).

Magnesium losses could be reduced by the second method, concentrating the mother liquor until $MgSO_4$ precipitates. Evaporation should not be allowed to proceed beyond the point where undesirable impurities such as NaCl precipitate with the $MgSO_4$ (MC-076). This procedure alone would not remove insoluble impurities such as silica, ferric oxide, aluminum oxide, and fly ash. After evaporation, however, the mother liquor can be sent to a deadend pond (as described above) or treated by conventional processes such as vapor compression distillation, reverse osmosis, or softening-ion exchange. These processes will be discussed in Section 6.7.

A possible complete treatment of the purge could be accomplished by method 3. The supernatant of soluble impurities could be treated by conventional water treatment techniques such as reverse osmosis, vapor compression distillation, flash evaporation, or softening-ion exchange. These processes will be discussed in Section 6.7. Pretreatment would be necessary before reverse osmosis or ion exchange processes. For reverse osmosis pretreatment is needed to reduce the concentration of calcium and magnesium to prevent scaling on the membrane. The concentration of impurities should be reduced before treatment by ion exchange to avoid frequent regeneration of the resin.

6.2 Double Alkali Sulfate Purge

In double alkali systems, some of the sulfer removed from the flue gas takes the form of soluble sodium sulfate because of oxidation in the system. Some of the active sodium (sodium associated with anions involved in SO_2 absorption reactions, including sulfite, bisulfite, hydroxide, and carbonate/bicarbonate) is thus converted to an inactive state which does not take part in SO_2 absorption. Converting Na_2SO_4 back to active sodium is relatively difficult. To do so, sulfate ion must be removed from the system while sodium is left in solution. Alternatively, the sodium sulfate may be removed from the system at the rate it is being formed. This alternative is not desirable since it wastes sodium. Furthermore, removal would generally be achieved by purging the sodium sulfate from the system in the liquor which is occluded in the wet solid waste product. The sulfate could then be leached from the waste. Water runoff can lead to contamination of surface water, and leaching and percolation of the leachate into the soil can result in contamination of the groundwater in the vicinity of the disposal site. Failure to allow for sulfate removal will ultimately result in: 1) precipitation of sodium sulfate in the system if active sodium (Na_2CO_3 or NaOH) is made up, or 2) eventual deterioration of the SO_2 removal capability due to loss of active sodium, if no makeup is added.

When low sulfur coal is burned, the ratio of oxygen to SO_2 is greater than it is when high sulfur coal is burned. Thus, a greater percentage of sulfite is oxidized to sulfate with low sulfur coal and a larger purge stream must be treated for sulfate removal.

Soluble nonsulfur/calcium species can also concentrate in the system and could be controlled by a purge. If the solid waste product is washed with excessive amounts of fresh makeup water to recover the potentially leachable sodium salts, the solubles can concentrate. Limited waste product washing provides these nonsulfur/calcium solubles with an exit through occlusion in the solid waste. As more sodium is recovered, more solubles are recovered. A purge may be necessary to maintain a desired level of concentration (KA-227).

6.2.1 Purge Characteristics

The sulfate purge characteristics are the same as those of the recirculated clarifier overflow. The purge contains sodium sulfate, sodium sulfite, and soluble nonsulfur/calcium species in varying amounts. The nonsulfur/calcium species enter the system in the lime and/or limestone, and in the makeup water. In systems having common prescrubber and absorber circulating loops, these species enter with the fly ash and flue gas. The soluble nonsulfur/calcium contaminant in highest concentration is probably sodium chloride. Contamination results from the absorption of HCl from the flue gas.

6.2.2 Effect on Receiving Stream

This stream could not be discharged directly to receiving streams due to the high levels of sodium, sulfate, sulfite, and other soluble species.

6.2.3 Treatment Technology

Several methods for sulfate removal in double alkali systems are discussed by Kaplan (KA-227). These methods are:

1) precipitation of sulfate as $CaSO_4 \cdot 2H_2O$ with the addition of lime (This method applies only to dilute double alkali systems);

2) co-precipitation of sulfate with calcium sulfite in a mixed crystal or solid solution;

3) addition of sulfuric acid;

4) formation of H_2SO_4 in an electrolytic cell;

5) limitation of oxidation.

With the first two methods, sulfates are removed in the regeneration operation and a sulfate purge is not required. With methods (3) and (4), the purge treated for sulfate removal can be returned to the system. With method (5), the purge is discharged with the solid waste.

If the active sodium concentration in the system is sufficiently low (less than about 0.15 M), sulfate can be removed from the system by precipitation as gypsum with the addition of lime, according to Equation 6.2-1:

$$Na_2SO_4 + Ca(OH)_2 + 2H_2O \rightarrow 2NaOH + CaSO_4 \cdot 2H_2O \qquad (6.2\text{-}1)$$

Because the solubility product of calcium sulfite is much lower than that of calcium sulfate, this reaction does not occur to a significant extent in the presence of high concentrations of sulfite ions. Thus, this method is not applicable for concentrated double alkali systems. This method for sulfate removal has been demonstrated in the dilute double alkali process developed by General Motors Corporation. This process has been applied to cleaning flue gas from coal-fired industrial boilers at the 30-40 MW (equivalent) level. The treated effluent is a clear solution of $NaOH/Na_2SO_3$ which is recycled to the absorber. The gypsum solid wastes are disposed of.

In concentrated systems, with a concentration of active alkali greater than about 0.15 M, sulfates cannot be removed to a significant degree by Equation 6.2-1. Under certain conditions, however, sulfate is co-precipitated with calcium sulfite in a mixed crystal or solid solution. This phenomenon has been described by Borgwardt as it applies to lime/limestone systems. Similar observations have been made by A. D. Little in their double alkali investigations. Enough sulfate can reportedly be removed by co-precipitation to allow successful operation without a sulfate purge with system oxidation rates as high as 20% of the SO_2 absorbed. The treated effluent is a clear solution of $NaOH/Na_2SO_3$ which is recycled back to the absorber. The solid wastes are disposed of.

In the third method (Equation 6.2-2), sulfuric acid is added to dissolve calcium sulfite, increasing the concentration of calcium ions in solution enough to exceed the solubility product of calcium sulfate.

$$Na_2SO_4 + 2CaSO_3 \cdot \tfrac{1}{2}H_2O + H_2SO_4 + 3H_2O \rightarrow 2NaHSO_3 + 2CaSO_4 \cdot 2H_2O \qquad (6.2\text{-}2)$$

This method requires more sulfuric acid than the stoichiometric amount indicated in Equation 6.2-2 since any material that functions as a base (such as unreacted lime or limestone) can consume sulfuric acid. Because of the high sulfuric acid consumption, this method may be uneconomical in applications with high oxidation rates. The sulfuric acid addition method is used for sulfate removal in full scale double alkali applications in Japan. It has also been pilot tested by A. D. Little in the U.S. The liquid effluent produced is a clear solution of sodium bisulfite which is recycled back to the main regeneration reactor. The gypsum solids are sold in Japan, but would probably be landfilled in the U.S.

The fourth method uses electrolytic cells developed by Ionics to remove sulfates as sulfuric acid and sodium hydroxide, according to Equation 6.2-3:

$$Na_2SO_4 + 3H_2O \xrightarrow[\text{Cell}]{\text{Electrolytic}} 2NaOH + H_2SO_4 + H_2 + \tfrac{1}{2}O_2 \qquad (6.2\text{-}3)$$

The technology for the cells has been relatively well developed by Ionics. In Japan, Kureha/Kawasaki have pilot tested the Yuasa/Ionics electrolytic process for sulfate removal with their double alkali process, and feel that it may be less expensive than the sulfuric acid addition method. Energy consumption for this method is relatively high, however. The sodium hydroxide solution produced can be recycled to the absorber. The sulfuric acid produced will be dilute. If no use for it can be found, it can be neutralized with lime and the resulting gypsum disposed of by landfill (KA-227, OT-R-051).

Another approach to sulfate control may be to limit oxidation by process and equipment design. If a sufficiently low level of oxidation can be maintained, it may be possible to remove sulfates by a small purge of sodium sulfate with the solid waste product. This alternative, however, would increase the potential for leaching from the solid waste (KA-227).

If a purge is necessary to maintain a desired level of soluble nonsulfur/calcium species, the remaining constituents of the double alkali purge can be removed with developed water treatment technology. Aerospace Corporation has reported that lime-soda softening reduces the concentrations of all constituents except soluble sodium and chloride salts in lime and limestone recirculating liquors (BO-203). Reverse osmosis can be used to further treat the water to remove the sodium chloride salts and other undesirable constituents. This results in a high quality product water stream with a low concentration of dissolved solids. Aerospace also reported that other existing treatment processes such as multistage flash evaporation, vapor compression distillation, and softening-ion exchange could be used instead of softening and reverse osmosis, but would generally be more expensive (BO-203). These treatment processes will be discussed in Section 6.7.

6.3 Prescrubber Blowdown

The Wellman-Lord Sulfite Scrubbing Process and the Magnesia Slurry Absorption Process require that chlorides and particulates are removed from the flue gases prior to entering the absorber. (With high chloride coals, a prescrubber may also be required for Double Alkali.) The prescrubber operates in a closed loop and requires a small blowdown to maintain desired levels of suspended and dissolved solids. An ESP is assumed to have removed 99 percent of the fly ash prior to entering the prescrubber. The remaining 1 percent of the fly ash and 100 percent of the chlorides are removed in the prescrubber. In the base case calculation (Section 4) the blowdown from the prescrubber was sized by the suspended solids limitation of a 5 percent solids slurry. The chlorides concentration was approximately 10,000 mg/ℓ (ppm).

Composition of the coal would have some effect on the quantity and composition of the blowdown stream. When coal with a high chloride content is burned, either the concentration of chlorides in the blowdown would be increased, or the amount of blowdown would be increased to keep the chloride concentration at a lower level. Coals with a high ash content may require an increased blowdown to maintain a low level of suspended solids, but since only 1 percent of the fly ash is assumed to be removed in the prescrubber, the change in quantity of the blowdown stream would be very slight.

6.3.1 Blowdown Characteristics

The prescrubber blowdown contains approximately 50,000 mg/ℓ (ppm) suspended solids, and 10,000 mg/ℓ (ppm) chlorides in the base case. Because many soluble inorganic salts can be leached from the fly ash, there are a wide variety of trace elements. These trace elements vary widely with coal type, and no typical analysis is practical. It can be assumed, however, that the constituents of the prescrubber blowdown stream will be equivalent to the constituents of the fly ash sluicing stream, with the significant addition of chlorides. The prescrubber blowdown is approximately 1 percent of the fly ash sluicing requirement.

6.3.2 Effect on Receiving Stream

The prescrubber blowdown stream is a wastewater source of high dissolved chloride concentration. The other components in the stream are of an equivalent nature, much reduced in quantity as compared to the fly ash sluice water. Thus, the major impact of this stream is the high chlorides concentration. The normal treatment procedure in current use is to route this stream to the ash pond to allow the suspended solids residence time to settle. The chlorides are diluted from 10,000 mg/ℓ to approximately 70 mg/ℓ (if an ash pond overflow of 2,000 gpm is assumed), and discharged to a receiving stream. The effect on receiving stream water quality is site specific and depends on the nature of the receiving stream water quality and flow parameters. Adverse impact is measured in relation to current regulations. These are set to maintain current water quality. The USGS Stream Classifications, given in Table 6.3-1, give an indication of the effect on the biosystems in the receiving streams.

TABLE 6.3-1. USGS STREAM CLASSIFICATIONS

Classification	Total Dissolved Solids mg/ℓ (ppm)
Fresh	0-1,000
Slightly Saline	1,000-3,000
Moderately Saline	3,000-10,000
Very Saline	10,000-35,000
Briny	>35,000

The impact of the ash pond overflow containing the diluted prescrubber blowdown is seen to be minimal, but certain receiving stream regulations may require treatment before discharge. In some systems, ash pond overflow is reused for ash sluicing instead of being discharged.

6.3.3 Treatment Technology

As the nation approaches a 1983 goal of zero discharge, recirculating systems will become predominant. In this case, the prescrubber blowdown stream will probably be treated separately, rather than sent to the ash pond. Currently available treatment technologies potentially applicable to this stream include reverse osmosis, vapor compression distillation, flash evaporation, and softening-ion exchange. These technologies will be discussed in Section 6.7. Before these technologies could be applied, however, the suspended solids content of the stream would have to be reduced, perhaps by sedimentation or filtration.

6.4 Cooling Water System Blowdown

The Wellman-Lord Sulfite Scrubbing Process, the Allied Chemical SO_2 Reduction Process, and the sulfuric acid production plant require cooling water. The most significant requirement is for condensation of evaporator overhead in the Wellman-Lord Process. The sulfuric acid plant requires cooling water for product acid cooling. The SO_2 gas stream in the Allied Chemical Process must be compressed, and cooling water is required for compressor seal cooling. This requirement is insignificant.

The circulating cooling water requirements, as defined in Section 4 for the application of FGD systems to the base case 500 MW power plant, are:

1)	Wellman-Lord condenser cooling water	1.3 m^3/s (20,000 gpm)
2)	Sulfuric acid product cooling, and	0.32 m^3/s (5,000 gpm)
3)	Compressor seal cooling, elemental sulfur process	0.019 m^3/s (300 gpm)

For comparative purposes, the circulating cooling water requirement for the condenser cooling system of the base case 500 MW power plant is 13 m^3/s (210,000 gpm).

6.4.1 Purge Characteristics

Cooling water systems can operate with once-through or recirculatory flows. Due to the nature of the once-through system, the chemical composition of the effluent water is essentially equivalent to that of the influent water. Recirculating cooling systems employ cooling devices such as cooling towers, spray ponds/canals, etc., which allow the reuse of cooling water. These devices promote cooling primarily by evaporating a portion of the recirculating water flow. Impurities and contaminants that come into the system with makeup water and other sources become concentrated in the system. A blowdown stream is withdrawn from the system to control the concentration of impurities and contaminants.

Soluble chemical species brought into the system by the makeup water are typically concentrated to levels ranging from 1,500 to 10,000 mg/ℓ (ppm). The chemical species contributing to the salinity of the blowdown is primarily determined by the makeup water. Chemical treatments to control corrosion, scale, biological fouling and solids deposition also impact the blowdown water quality. The intimate contact between air and water in the cooling devices enables particulate matter and soluble gases to be scrubbed from the air. Up to 80 percent of the suspended solids in recirculating systems are estimated to have originally come into the system as airborne particulates (GL-028).

6.4.2 Effect on Receiving Streams

The quality of the FGD-associated cooling water stream is similar to that of the power plant cooling water system. The FGD-associated cooling water requirements are very small compared to power plant cooling requirements (less than 10%). The blowdown from the FGD cooling systems could thus be treated and discharged along with the power plant cooling water blowdown, and would not have a significant impact on receiving stream water quality.

6.4.3 Treatment Technology

Cooling water system blowdown is normally sent to a hold pond to allow residence time for suspended solids settling. In current practice, the supernatant is then discharged to a receiving stream without further treatment for removal of dissolved solids. As the nation approaches the 1983 goal of zero discharge, recirculatory systems will predominate. Such treatment technologies as reverse osmosis, vapor compression distillation, flash evaporation, and softening-ion exchange may be used to treat small blowdowns for dissolved solids removed. These treatment technologies are discussed in Section 6.7.

6.5 Possible Lime/Limestone Purge

Under normal operating conditions no lime/limestone system purge should be necessary. Normal operating conditions are defined by design criteria, and include characteristics of the boiler, coal, hardware, absorbent, and disposal facilities. Under normal conditions, water is lost by evaporation in the scrubber, and occlusion in the solid waste. Normally, sufficient dissolved solids are occluded in the solid waste to maintain recirculated dissolved solids at desired concentrations.

In a recent study, Aerospace defined four situations which might require a purge of scrubbing liquor to allow sufficient makeup of fresh water to avoid scaling (BO-203). These include:

1) operation below some critical flue gas mass load level (created by low boiler loading);

2) use of lower sulfur coal than that for which the system was designed;

3) oxidation of sulfite sludge to a sulfate sludge, with more liquor recovered due to greater dewatering efficiency; and

4) need for a quick blowdown due to catastrophic circumstances.

Aerospace estimated that a purge might become necessary in the range of 50 percent of the maximum design criteria. During start-up or shutdown, a temporary excess liquor problem can usually be handled by the excess capacity of the disposal pond. Operator error can also result in a purge stream.

6.5.1 Purge Characteristics

The possible purge from the lime/limestone system would have the same quality as the recirculated clarifier supernatant. This liquor differs from system to system due to coal type, fly ash collection facilities, and scrubber operation.

The liquor is saturated with dissolved calcium sulfite and/or calcium sulfate salts. Sodium and chloride ions are also present in high concentrations. Most trace elements and toxic species are controlled below 1 mg/ℓ (ppm) (BO-203). Table 6.5-1 shows the range of constituents in four scrubber liquors sampled by Aerospace Corporation.

6.5.2 Effect on Receiving Streams

Aerospace Corporation has concluded that purged scrubber liquor must be treated if it is to be discharged to a receiving stream (BO-203). This statement was based on data from "Disposal of By-Products from Nonregenerable Flue Gas Desulfurization Systems", which is a study conducted under EPA Contract 68-02-1010. A secondary reason for not allowing direct discharge of these liquors is the Federal Water Pollution Control Act Amendments of 1972. Requirements for 1977 and 1983 discharge limitations, based on best practical control technology available and the best available technology economically available, also prohibit direct discharge.

6.5.3 Treatment Technology

Aerospace Corporation has studied the feasibility of using available and developmental water treatment technologies applicable to scrubber liquors. They have concluded that lime-soda softening would render the scrubbing liquors acceptable for power plant general services use, but the liquor would still contain a high concentration of dissolved solids. Lime-soda softening followed by reverse osmosis would make the liquor suitable for public water supply. The softening step must precede treatment by reverse osmosis because the high concentration of calcium in the purged liquor would cause scaling of the membrane. Other

TABLE 6.5-1. RANGE OF CONCENTRATION OF CONSTITUENTS IN SCRUBBER LIQUORS STUDIED

Constituents	Range of constituent concentrations at potential discharge points[a] (mg/ℓ)
Aluminum	0.03 - 0.3
Antimony	0.09 - 2.3
Arsenic	<0.004 - 0.3
Beryllium	<0.002 - 0.14
Boron	8.0 - 46.
Cadmium	0.004 - 0.11
Calcium	520. - 3000.
Chromium (total)	0.01 - 0.5
Cobalt	0.10 - 0.7
Copper	<0.002 - 0.2
Iron	0.02 - 8.1
Lead	0.01 - 0.4
Magnesium	3.0 - 2750.
Manganese	0.09 - 2.5
Mercury	0.0004 - 0.07
Molybdenium	0.91 - 6.3
Nickel	0.05 - 1.5
Potassium	5.9 - 32.
Selenium	<0.001 - 2.2
Silicon	0.2 - 3.3
Silver	0.005 - 0.6
Sodium	14.0 - 2400.
Tin	3.1 - 3.5
Vanadium	<0.001 - 0.67
Zinc	0.01 - 0.35
Carbonate	<1.0 - <10.
Chloride	420. - 4800.
Fluoride	0.07 - 10.
Sulfite	0.8 - 3500.
Sulfate	720. - 10,000.
Phosphate	0.03 - 0.41
Nitrogen (total)	<0.001 - 0.002
Chemical oxygen demand	60. - 390.
Total dissolved solids	3200. - 15,000[b]
Total alkalinity (as $CaCO_3$)	41. - 150.
Conductance mho/cm	0.003 - 0.015
Turbidity, Jackson Units	<3. - <10.
pH	3.04 - 10.7

[a] Samples obtained from: EPA/TVA, Shawnee, Steam Plant - venturi and spray tower; EPA/TVA Shawnee Steam Plant - turbulent contact absorber; Arizona Public Service Cholla Station - flooded disk scrubber and absorption tower; and Duquesne Light Phillips Station - single- and dual-stage venturi.

[b] Includes all soluble species.

existing treatment processes such as flash evaporation, vapor compression distillation, and softening-ion exchange could be used instead of lime-soda softening and reverse osmosis, but were not economically competitive. These treatment processes are discussed in Section 6.7.

6.6 Lime/Limestone/Double Alkali Solid Waste

The solid waste produced by the nonregenerable FGD processes does not directly impact receiving stream water quality. However, the solid waste can impact surrounding water quality through leaching and percolation of soluble components of the solid waste. This subject is discussed further in the report on solid waste impact (RO-359).

6.7 Existing Water Treatment Technologies Applicable to Wastewater from FGD Systems

Several existing technologies for water treatment have been identified as potentially being applicable for treatment of water from FGD processes. These technologies will be discussed in this section.

6.7.1 Lime-Soda Softening

This process can be used to decrease the concentration of calcium and magnesium ions in purge liquors from the lime/limestone, magnesia slurry, and double alkali processes. It can also be used as a pretreatment step before reverse osmosis or ion exchange processes. Lime (CaO) and soda ash (Na_2CO_3) added to the liquor react with the major species in the liquor to precipitate the calcium and magnesium together with the heavy metals. Carbon dioxide is added to adjust the pH, and the solution is filtered and centrifuged. The precipitates are sent to the disposal site (BO-203). Concentrations of hardness (magnesium and calcium ions) in the product water can be reduced to 50 mg/ℓ (expressed as $CaCO_3$) (WE-324). An 80 to 90 percent reduction in the concentrations of As, Cd, Mn, Pb, and Se and a 30 to 90 percent reduction in hexavalent chromium can also reportedly be achieved with this process. Boron will also be removed to some extent (BO-203). Other dissolved solids such as sodium and chlorides will not be removed, however. Thus, the product water may have to be further treated by reverse osmosis or ion exchange before it can be recycled to the process. Lime-soda softening processes are in commercial operation.

6.7.2 Reverse Osmosis

Reverse osmosis could be used to decrease the concentration of dissolved solids in the prescrubber blowdown from the Magnesia Slurry or Wellman-Lord; cooling tower blowdown from the processes requiring it; or purge streams from the Magnesia Slurry, Double Alkali, or Lime/Limestone Processes. Because the waste stream produced from reverse osmosis would be large in volume (around 25 percent of the feedwater), another process, such as vapor compression distillation, would have to be used in conjunction with reverse osmosis to treat this waste stream. The purge streams would contain relatively high concentrations of calcium and/or magnesium ions and would have to be treated (perhaps by lime-soda softening) to decrease these concentrations before the streams could be treated by reverse osmosis. Otherwise, the calcium or magnesium salts could cause membrane fouling.

In reverse osmosis, water is forced through a semipermeable membrane which allows the passage of water but prevents the passage of dissolved solids and other impurities. The resulting two products are deionized water and brine concentrate. The net driving force for flow of water through the membrane is the total applied pressure less the osmotic pressure of the system. There are many inorganic and synthetic organic materials that possess the property of semipermeability, but cellulose acetate is the most common membrane material employed (GE-009).

Reverse osmosis membrances are subject to fouling from many different fouling agents. The most significant of these agents are:

- biological growths
- suspended solids or particulate matter
- scale
- manganese and iron
- organics

To control membrane fouling, a feedwater pretreatment step is often dictated and certain limitations are imposed on the process. Process feed pretreatment can take several forms depending upon each particular feedwater.

In addition to being a pretreatment consideration, scale control also imposes restrictions or limitations on the reverse osmosis process. Scale formation is driven by supersaturation of a solution with respect to a chemical salt. Calcium carbonate [$CaCO_3$], calcium sulfate [$CaSO_4$], and magnesium hydroxide [$Mg(OH)_2$] are the primary scaling salts due to their low solubilities. Therefore, these salts must be kept below their saturation concentrations in order to avoid precipitation and scale formation. The conversion of saline water to desalted water by reverse osmosis is thus limited in the degree of feedwater concentration that can be achieved. The degree of concentration varies with the quality of feedwater to the process.

A high degree of concentration may not be possible for wastes from FGD purge streams because of the high concentrations of calcium and/or magnesium.

Reverse osmosis units have typically demonstrated 90-98% salt rejection for feedwater with about 10,000 ppm dissolved solids (comparable to the concentration in prescrubber blowdown). Product water TDS levels of less than 200 ppm can be achieved with one stage, with 75% recovery of the feedwater. Multiple stage RO systems can be used to achieve lower product TDS levels (<50 ppm). Reverse osmosis will also remove more than 95% of organics and all colloidal particles down to 0.05 microns (DI-149). The concentrated waste stream from a reverse osmosis unit will be relatively large in volume. The actual volume will depend on the amount of feedwater recovered. This waste stream can be further concentrated by vapor compression distillation or, in some applications, sent to an evaporation pond.

Reverse osmosis units are commercially marketed by a number of companies. They have been used to treat cooling tower blowdown water to recover deionized water for reuse, and to produce drinking water from sea water and inland brackish water.

6.7.3 Ion Exchange

Ion exchange is a commercially available water treatment process that has been suggested as a possible means for treating wastewater for FGD processes (BO-203). Because of the high concentration of dissolved solids in these streams, however, treatment by ion exchange may be prohibitively expensive. It has been estimated that processes such as reverse osmosis would be more economical than ion exchange for treating water with an inlet TDS concentration of 1000 mg/ℓ (ppm) or more (WE-324).

The wastewater from FGD would generally have more than 10,000 mg/ℓ (ppm) of TDS.

Ion exchange is based on the reversible interchange of ions between a solid and liquid phase in which there is no permanent change in the structure of the solid (ST-135). As water containing soluble impurities flows across a bed of ion exchange resin, ions in the resin are replaced with different ions from the water. When the resin has reached its load capacity, it is regenerated by rinsing with a regenerant solution that contains a high concentration of the type of ions contained in the original resin. When water with a high concentration of TDS is treated by ion exchange, a large quantity of impurities is removed and frequent flushing and regeneration of the beds is required. Wastewater produced from ion exchange includes backwash water, spent regenerant solution, and rinse water.

6.7.4 Vapor Compression Distillation

Vapor compression distillation can be used to further concentrate the waste stream from a reverse osmosis unit. It can also treat the prescrubber blowdown from the Wellman-Lord or Magnesia Slurry Processes; cooling tower blowdown from the processes requiring it; or purge streams from the Magnesia Slurry, Double Alkali, or Lime/Limestone Processes.

Vapor compression distillation (also known as "brine concentration") uses a falling film evaporator along with a vapor compression thermodynamic cycle to concentrate the TDS in waste streams while producing low TDS product water. The pH of the waste stream to be treated is adjusted to between 5.5 and 6.5 for decarbonation. The stream is then heated by heat exchange with the hot product condensate, deaerated, and combined with the slurry concentrate in the evaporator sump. The brine slurry is constantly circulated from the sump to the top of the evaporator tubes. As the slurry falls as a film on the inside of the tubes, part of the water is vaporized by steam condensing on the outside of the tubes. This vapor is compressed and introduced to the shell side of the evaporator. The steam condenses on the outside of the tubes, and the condensate is pumped through the feed preheater. This product condensate contains less than 10 mg/ℓ (ppm) TDS, and is suitable for reuse in the power plant or FGD process. Approximately 90% of the inlet water can be recovered when treating water with an inlet TDS of 10,000 mg/ℓ (ppm) (comparable to the water quality of the prescrubber blowdown) (RE-259). The brine concentrate, which will be about 10% of the original stream volume and contain most of the dissolved solids, can be sent to an evaporation pond or mechanical drying system for final disposal. Vapor compression distillation processes are commercially available. They have been installed in electric power generating stations in the western and southwestern states to recover deionized water from cooling tower blowdown (LE-299). Vapor compression distillation is an energy intensive process requiring approximately 90 kw-hr/1000 gal of water processed (RE-259). Most of this energy goes into driving the vapor compressor.

6.7.5 Multistage Flash Evaporation

Multistage flash evaporation could be used to treat the prescrubber blowdown from the Wellman-Lord or Magnesia Slurry Processes; cooling tower blowdown from any of the processes; or purge streams from the Magnesia Slurry, Double Alkali, or Lime/Limestone Processes. In this method, the wastewater stream is heated and flash evaporated in a series of stages under progressively lower pressures. Each stage of the evaporator has a heat exchanger for preheating the incoming liquid and a vacuum chamber for flash evaporation. The vapor is used for preheating the incoming liquid, and is condensed. The condensate from each stage is collected. This product water will have a low concentration of TDS (less than 50 ppm), and is suitable for reuse in the system (even as boiler feedwater) or for discharge. A portion of the residual concentrated wastewater is mixed with the influent stream for recycle through the evaporator. The remaining concentrated waste can be sent to an evaporation pond or mechanical drying system for final disposal. The amount of waste produced will depend on the evaporator design and equipment operating characteristics. For example, if water with an inlet TDS of 10,000 ppm (comparable to prescrubber blowdown or lime/limestone purge) is treated to a product concentration of less than 50 ppm TDS and a waste concentration of about 100,000 ppm TDS, the waste stream will be about 10 percent of the feedwater. Multistage evaporators have been used in the chemical process industry for many years, and have also been used for desalting sea water to produce drinking water (BO-203).

REFERENCES

AY-007 Aynsley, Eric and Meryl R. Jackson, *Industrial Waste Studies: Steam Generating Plants*, *Draft Final Report*. Rosemont, Ill., Freeman Labs., Inc., 1971.

BA-185 Babcock & Wilcox, *Steam/Its Generation and Use*. 38th ed. New York, 1972.

BE-162 Bell, William E. & E. Dennis Escher, "Disposal of Chemical Cleaning Waste Solvents". *Mat. Protect. Perfor.* 9, 15 (1970).

BL-036 Blake, R. T., "Proper Feedwater Treatment Helps Minimize Pollution", *Plant Eng.* 1970 (Oct), 34.

BO-203 Bornstein, L. J., et al., *Reuse of Power Plant Desulfurization Waste Water*, *Final Report. EPA-600/2-76-024, PB250732.* El Segundo, California, Aerospace Corp., Feb. 1976.

CH-R-387 Christman, Peter G., *Water Recycle/Resuse Alternatives at the (Pennsylvania Power and Light) Montour Station*, *Draft Report. DCN 77-200-118-02, EPA, Contract No.* 68-03-2339, TN 200-118-11. Austin, TX, Radian Corp., April 1977.

CL-028 Cline, M. A. A., B. A. Thrush, and R. P. Wayne, "Kinetics of the Chemiluminescent Reaction Between NO and O_3", *Trans. Faraday Soc.* *60*(494), 359-70 (1964).

CO-380 "Coal Preparation and Unit-Train Loading", *Coal Age* *1970* (July), 188.

CU-077 Cuffe, Stanley T., *Alternative Control Systems to be Considered for Cost and Environmental Impact Analysis-Program to Review the NSPS for Emissions from Steam Generators*, Research Triangle Park, N.C. EPA, April 1977.

DA-189 Davis, John C., "Coal Cleaning Readies for Wider Sulfur-Removal Role", Chem. Eng. 83 (5), 70 (1976).

DE-064 Deurbrouck, A. W., *Sulfur Reduction Potential of the Coals of the United States*. Report of Investigations 7633. Pittsburgh, Pa., Pittsburgh Energy Research Center, 1972.

REFERENCES CONTINUED

DI-R-139 Dickerman, J. C., et al., *An Investigation of the Use of Coal in the HL&P Generating System*, Interim Report, 2 vols. DCN 77-200-162-01. Austin, TX, Radian Corp., Feb. 1977.

DO-048 Donohue, John M., "Making Cooling Water Safe for Steel and Fish, Too", *Chem. Eng. 78*(22), 98 (1971).

DO-051 Donahue, John M., "Chemical Treatment", *Ind. Water* Eng. 7(5), 35 (1970).

EN-127 Environmental Protection Agency, (Office of Air and Water Programs, Effluent Guidelines Div.), *Development Document for Proposed Effluent Limitations Guidelines and New Source Performance Standards for the Steam Electric Power Generating Point Source Category*. 1974.

FI-102 "Fine-Coal Treatment and Water Handling", *Coal Age 66* (12), 67 (1961).

GA-R-203 Gathman, Wayne and J. G. Noblett, Jr., *Water Recycle/Reuse Alternatives at the Public Service of Colorado Comanche Plant*. Tech. Note 200-118-09, EPA Contract No. 68-03-2339. Austin, TX, Radian Corp., Aug. 1976.

GE-009 Gentry, Robert E., Jr., "Reverse Osmosis: A Pleasant Inversion", *Env. Sci. Tech. 1*, 124-31 (1967).

GL-028 Glover, G. E., "Cooling Tower Blowdown Treatment Costs", *Industrial Process Design for Water Pollution Control*, Vol. 2. New York, AICHE, pp. 74 ff.

HA-R-636 Hargrove, O. W. and J. G. Noblett, Jr., *Water Recycle/Reuse Alternatives at the Arizona Public Service Four Corners Stations*. Technical Note 200-118-07, EPA Contract No. 68-03-2339, Reviewed by D. M. Ottmers, Jr., Austin, TX, Radian Corp., July 1976.

HU-051 Hunter, William D., Jr., "Application of SO_2 Reduction in Stack Gas Desulfurization", Presented at the Flue Gas Desulfurization Symposium, New Orleans. May 1973.

IF-001 Ifeadi, C. N. and H. S. Rosenberg, "Lime/Limestone Sludge Disposal Trends in the Utility Industry", Presented at the Flue Gas Desulfurization Symposium, Atlanta, Ga., Nov. 1974.

REFERENCES CONTINUED

KA-227 Kaplan, Norman, "Introduction to Double Alkali Flue Gas Desulfurization Technology", Presented at the EPA Flue Gas Desulfurization Symposium, New Orleans, Louisiana. March 1976.

LE-218 Leonard, Joseph W. and David R. Mitchell, eds., Coal Preparation, 3rd edition. New York, The American Institute of Mining, Metallurgical and Petroleum Engineers, Inc., 1968.

LO-071 Lowry, H. H. ed., Chemistry of Coal Utilization, 2 vols. & supplementary volume. N.Y., Wiley, 1945, 1965 (supplementary volume).

MA-411 Martin, John F., "Coal Refuse Disposal in the Eastern United States". News Env. Res. Cincinnati. Dec. 1974.

MC-076 McGlamery, G. G., et al., Conceptual Design and Cost Study. Sulfur Oxide Removal from Power Plant Stack Gas. Magnesia Scrubbing-Regeneration: Production of Concentrated Sulfuric Acid. EPA-R2-73-244. Muscle Shoals, Ala., TVA, 1973.

MC-147 McGlamery, G. G., et al., Detailed Cost Estimates for Advanced Effluent Desulfurization Processes, Final Report. EPA-600/2-75-006. Muscle Shoals, Ala., TVA, Jan. 1975

MA-230 Marshall, Wm. L., "Cooling Water Treatment in Power Plants", Ind. Water Eng. 9(2), 38 (1972).

NO-R-106 Noblett, James G., Jr., Water Recycle/Reuse Alternatives at the Georgia Power Company Plant Bowen. Technical Note 200-118-08, Radian Project No. 200-118, EPA Contract No. 68-03-2339. Austin, TX., Radian Corp., Aug. 1976.

NO-R-137 Noblett, James G., Jr., Water Recycle/Reuse Alternatives at Montana Power Company's Colstrip Plant. DCN 77-200-188-04, Technical Note 200-118-12, EPA Contract No. 68-03-2339, Reviewed by D. M. Ottmers. Austin, TX., Radian Corp., May 1977.

OT-R-051 Ottmers, D. M., Jr., et al., Evaluation of Regenerable Flue Gas Desulfurization Processes, revised report, 2 vols., EPRI RP 535-1, Austin, TX., Radian Corp., July 1976.

REFERENCES CONTINUED

PE-030 Perry, John H., *Chemical Engineers Handbook*, 4th ed. New York, McGraw-Hill, 1963.

PE-259 PEDCo Environmental, Inc., *Flue Gas Desulfurization Systems*, Jan. Feb., March, 1977, summary report, EPA Contract No. 68-02-1321, Task No. 28, Cincinnati, Ohio, 1977.

RA-R-352 Radian Corporation, *An Investigation of the Potential for Utilization of Saline Ground Water in Energy-Related Processes*, Draft Final Report. DCN 77-200-152-01 (06), FE-200-1, ERDA Contract No. E(49-18)-2000. Austin, TX., Jan. 1977.

RO-359 Rossoff, J. and P. P. Leo, *The Solid Waste Impact of Controlling SO_2 Emissions from Coal-Fired Steam Generators*. El Sungundo, CA., Aerospace Corp., Oct. 1977.

RI-160 Rice, James K. and Sheldon D. Strauss, "Water-Pollution Control in Steam Plants", *Power* *120*(4), S1-20 (1977).

ST-135 Strauss, Sheldon D., "Water Treatment", *Power* *117*(6), SI-S24.

WE-003 (Paul) Weir Company, *An Economic Feasibility Study of Coal Desulfurization*, 2 vols. PB 176 845, PB 176 846. Chicago, Ill., Oct. 1965.

WE-324 Weber, Walter J., Jr., *Physiochemical Processes for Water Quality Control*, New York, Wiley, 1972.

APPENDIX A

A.0 EXPLANATION OF CALCULATIONAL METHOD

This section presents example calculations, and discussions of the assumptions used in estimating the water consumption of uncontrolled power plants, and SO_x control strategies.

A.1 Selection of Representative Coal Ultimate Analyses

The model plant systems require the selection or approximation of representative ultimate analyses for the following four coals:

coal #1) 0.8% S (dry basis), 8,000 Btu/lb (AR basis), 6% ash, 30% H_2O
coal #2) 0.8% S (dry basis), 11,000 Btu/lb (AR basis), 6% ash, 15% H_2O
coal #3) 3.5% S (dry basis), 12,000 Btu/lb (AR basis), 12% ash, 2.6% H_2O
coal #4) 7.0% S (dry basis), 12,000 Btu/lb (AR basis), 12% ash, 5.7% H_2O

Also, the determination of the ultimate analyses following coal cleaning operations, where 40% of the sulfur is to be removed, is required for the following two original coals:

coal #5) 3.5% S (dry basis), 12,000 Btu/lb (AR basis), 12% ash, 2.6% H_2O
coal #6) 7.0% S (dry basis), 12,000 Btu/lb (AR basis), 12% ash, 5.7% H_2O

Table B.1-1 shows the ultimate analyses of several example coals.

Coal #1 - The ultimate analyses of coal #1 was approximated by taking midrange values of the three low heating values western coals listed in Table B.1-1, i.e., coals 15, 16, and 17.

Coal #2 - Coal 14 is the only high heating value western coal listed in Table B.1-1. From a comparison of proximate analyses, this coal seems to have a representative ultimate analysis. Therefore, the ultimate analysis of coal #2 is taken to be the same as coal 14.

TABLE A.1-1. EXAMPLE COAL ULTIMATE ANALYSES

No.*	Type	C	H_2	N_2	S	O_2	Ash	H_2O	Btu/lb
1	PA	79.84	1.78	0.25	0.71	1.96	9.7	4.5	12,745
2	PA	79.45	2.21	0.77	0.60	1.95	11.9	2.5	12,925
3	VA	70.00	3.24	0.77	0.62	2.55	20.2	2.0	11,925
4	WV	84.21	4.47	1.21	0.74	2.51	5.1	1.0	14,715
5	PA	77.52	4.16	1.30	1.68	2.08	10.3	1.3	13,800
6	PA	76.74	4.15	1.38	1.68	2.68	10.2	1.5	13,720
7	PA	75.42	4.48	1.21	2.20	2.84	10.2	1.5	13,800
8	PA	72.66	4.62	1.45	1.82	4.96	11.2	1.5	13,325
9	KY	79.94	5.14	1.50	0.70	6.26	3.3	2.5	14,480
10	OH	67.39	4.75	1.17	4.00	6.16	9.1	3.6	12,850
11	IL	64.24	4.39	1.28	2.70	7.26	11.7	5.8	11,910
12	UT	69.83	4.90	1.49	0.90	10.33	6.4	5.2	12,600
13	IL	59.88	4.31	1.13	3.20	7.18	9.0	12.2	11,340
14	MT	63.48	4.00	1.02	0.43	9.57	7.0	14.1	11,140
15	WY	53.89	3.62	1.14	0.30	12.07	3.7	25.0	9,345
16	WY	47.10	3.56	0.57	0.55	11.83	4.8	31.0	8,320
17	ND	42.46	2,86	0.53	0.40	12.15	4.2	37.0	7,255
18		70.7	4.7	1.1	3.4	10.3	7.1	2.7	12,400
19		53.13	3.70	1.00	0.39	14.17	4.62	23	
20	Western	72.7	5.3	1.1	1.0	9.0	8.9	2.0	13,135
21	Eastern	69.9	4.9	1.3	1.1	7.1	13.7	2.0	12,640

* Coals 1 → 17 are from Steam

Coals 18 & 19 are from Tech. Note 200-118-09

Coals 20 + 21 are from FA-084

Coal #3 - The ultimate analysis used for coal #3 is the same as that used by McGlammery, et al (MC-147). A comparison of this analysis with the analyses of coals 7, 10, 11, and 13 in Table A.1-1 shows similar values.

Coal #4 - Coal 10 is the only high sulfur coal listed in Table B.1-1, and is 4% sulfur rather than 7% sulfur. Values of 12,000 Btu/lb, 5.7% H_2O, and 12% ash were assumed. The remaining properties of coal #4 were chosen to be similar to those of coal 10.

Coals #5 and #6 are the product coals from washing Coals #3 and #4. It is assumed that 1) 40% of the original sulfur content has been removed, 2) 50% of the original ash has been removed (RA-215), 3) the yield is 85% (ZI-014, MA-495, RA-215), and 4) that the coal product is dewatered to 15% (FI-102, LE-218, CO-380).

The ultimate analyses of the six representative coals, chosen for use in our study calculations, are shown in Table A.1-2.

TABLE A.1-2. SIX REPRESENTATIVE COALS - ULTIMATE ANALYSIS

	Ultimate Analysis	Coal No. 1	2	3	4	5	6
	H_2O	30.0	15.0	2.6	5.7	15.0	15.0
	C	47.3	63.8	70.5	63.6	66.6	63.1
	H_2	3.5	4.0	4.5	5.0	4.3	5.0
	S	0.6	0.7	3.4	6.6	1.9	4.0
	N_2	0.6	1.0	1.0	1.1	0.9	1.1
	O_2	12.0	9.5	6.0	6.0	5.7	5.9
	Ash	6.0	6.0	12.0	12.0	5.6	5.9
Average Heating Value	Btu/lb	8,000	11,000	12,000	12,000	11,100	11,500

A.2 Base Case Assumptions

A set of design criteria constituting base case was chosen to facilitate discussions and comparisons in the body of the report. Also, in Appendix B, this base case is used to illustrate the detailed calculational procedures. This section defines the base case.

A.2.1

The following assumptions are the same as those used by McGlammery, et al to calculate detailed mass balances for FGD systems (MC-147):

500 MW power plant
37% conversion efficiency
3.5% sulfur coal (dry basis)
12% ash (as fired basis)
12,000 Btu/lb (as fired basis)
92% of sulfur in coal evolves as SO_2
75% of ash evolves as fly ash
20% excess air
13% leakage
90% SO_2 removal

In addition, the base case that we have used assumes that:

0.13 lb H_2O/lb dry air - humidity of excess air
45% of total heat is lost to condenser cooling system

20°F cooling water temperature used in the condenser

99% of fly ash removed in an ESP prior to entering the scrubbing system

5% sluicing of fly ash

1% sluicing of bottom ash

The assumptions applicable to the FGD systems studied are:

Lime: Adiabatic saturation of the flue gases in the scrubber

105% stoichiometric lime

pond evaporation equals rainfall

first stage S/L separator produces a 40% solids sludge

ponding or second stage S/L separator produces a 60% solids sludge

10% solids recirculating slurry

25% oxidation

Limestone:

adiabatic saturation of the flue gases in the scrubber

120% stoichiometric limestone

pond evaporation equals rainfall

25% oxidation

10% solids recirculating slurry

first stage S/L separator produces a 40% solids sludge

ponding or second stage S/L separator produces a 60% solids sludge

Wellman-Lord:

- 15% of circulating slurry is sent to purge treatment
- double effect evaporators
- 60% overhead in each effect
- 10% H_2O in product SO_2 stream

Magnesia Slurry:

- 105% stoichiometric MgO
- 10% solids recirculating slurry
- first stage S/L separator produces a 40% solids sludge
- thermal decomposition of $MgSO_3 \cdot 6H_2O$
- 95% solids cake from centrifuge

Double Alkali:

- 105% stoichiometric lime
- 120% stoichiometric limestone
- pond evaporation equals rainfall
- first stage S/L separator produces a 40% solids sludge
- second stage S/L separator or ponding produces a 60% solids sludge
- 25% oxidation

A.3 Calculation of Flue Gas Rates and Compositions

To calculate flue gas rates and compositions, a coal ultimate analysis was converted to a molal composition, and the amount of excess air calculated. Then the volumes of combustion products were calculated based on 100 lb of fuel fired. A coal usage rate was then calculated, and from this the total flue gas volume was calculated. The following calculations for the base case illustrate the calculational method used for all the model systems.

A.3.1 Base Case Calculations

Basis: 100 lb coal

	Ultimate Analysis	Mol. Wt.	Moles
H_2O	2.6	18	0.14
C	70.5	12	5.88
H_2	4.5	2	2.25
S	3.4	32	0.11
N_2	1.0	28	0.04
O_2	6.0	32	0.19
Ash	12.0		

For 100% combustion:

	Moles	Moles O_2	Moles Air
C	5.88	5.88	
H_2	2.25	1.13	
S	0.11	0.11	
		7.12	
O_2 in coal		0.19	
		6.93	33.00

With 20% excess air and 13% leakage:

Air = 33.00 (1.33) = 43.89 moles

O_2 = 9.22 moles

H_2O in excess air = (43.89) (0.013) ($^{29}/18$) = 0.92 moles

Excess air = 10.89 moles

Excess O_2 = 2.29 moles

Combustion products:

	Moles	Mol. Wt.	lb
CO_2	5.88	44	285.72
H_2O	3.31	18	59.58
SO_2	0.10	64	6.40
N_2	35.71	28	971.88
O_2	2.29	32	73.28
	46.29		1,369.86

Avg. Mol. Wt. = 29.6

Flue gas volume:

$$V = \frac{NRT}{P} = \frac{(46.29 \text{ moles})}{(14.7 \text{ psia})} \left(10.73 \frac{\text{psia ft}^3}{\text{mole °R}}\right) (810\text{°R})$$

$$= 27,369 \text{ acf}$$

$$= 16,624 \text{ scf/100 lb coal fired}$$

Coal usage rate:

500 MW @ 37% eff = 4.492 x 10^9 Btu/hr

Coal AHV = 12,000 Btu/lb

Usage rate = 374,342 lb/hr

Flue gas rate:

$$(374.342 \text{ lb/hr}) (166.24 \text{ scf/lb}) \left(\frac{1 \text{ hr}}{60 \text{ min}}\right)$$
$$= 1{,}037{,}177 \text{ scfm}$$

For one of four equivalent scrubbing trains the flue gas rate would be: 259,000 scfm.

A.3.2 Model Systems Results

Each type of coal will require a characteristic coal usage rate, and therefore the flue gases resulting from coal combustion will have differing compositions and rates. In this study, it is assumed that the conversion efficiency does not vary with power plant size. Therefore, to calculate flue gas rates for the model systems it is necessary to make separate calculations for each coal type, but power plant size and flue gas rate are directly proportional and thus results can be scaled. Table A.3-1 shows the flue gas rates and compositions for each of the example coals for a 500 MW power plant.

TABLE A.3-1. FLUE GAS RATES AND COMPOSITIONS FOR EXAMPLE COALS (500 Mw plant; one of four trains)

Flue Gas Composition	Coal No.*					
	1	2	3	4	5	6
Gas Rate**	307.2	265.9	259.3	254.4	282.5	271.9
CO_2	12.41	12.91	12.70	12.32	12.56	11.94
H_2O	12.63	8.81	7.15	5.67	8.69	9.51
SO_2	0.06	0.05	0.22	0.46	0.14	0.27
N_2	70.27	73.41	74.98	76.50	73.76	73.45
O_2	4.63	4.83	4.95	5.04	4.86	4.83

*All composition values are in volume percent

**Gas rates are given for 500 MW plants; given in 1000 scfm for one of four trains

A.4 Water Requirements For A 500 MW Power Plant

The processes and operations in a coal-fired power plant that require water consumption are:

- cooling water
- ash handling
- boiler blowdown
- water conditioning
- equipment cleaning, and
- miscellaneous.

For the purpose of calculating a power plant water balance for an estimation of the impact that an FGD installation would have on power plant water consumption, a 500 MW power plant was chosen for comparative basis. Due to general unavailability of data on intermittant and miscellaneous water consumptive practices, water conditioning, equipment cleaning, and miscellaneous operations will be addressed as "general service water", and plant data from preliminary water recycle/reuse studies currently being conducted by Radian under EPA contract will be used to characterize these flows.

Current water management in the power generation industry uses two basic techniques: 1) once-through, and 2) recirculatory. There is a wide variability in plant layouts, and many combinations of once-through, and recirculatory practices are used in various power plant water systems. The two extremes in water management would be where: 1) once-through techniques are used in all systems, and 2) recirculatory practices are maximized in all systems. The total recirculatory system ultimately results in zero discharge. The once-through system requires enormous quantities of water. Current practice falls somewhere between these two extremes, but because of the large number of available process options and possible combinations, there is no "typical" water system. Water balances will be calculated for power plant water systems that operate 1) once-through, 2) with recirculatory cooling at 2.5 cycles of concentration, and once-through ash handling, 3) with recirculatory cooling at 5 cycles with 50% recycle ash handling and 4) with zero discharge of concentration, and once-through ash handling. These four cases should characterize representative water consumption in the current coal-fired power generation industry. Hereafter, they will be referred to as cases #1 through #4.

A.4.1 Cooling Water System

Assume: 1) "45% of a fossil-fuel fired generating station's energy is removed and ultimately discharged to the environment by the condenser cooling system" (DI-139)

2) power plant conversion efficiency is 37% (MC-147)

∴ For a 500 MW power plant 1,351 MW of energy are produced from the coal burned

$$(1{,}351\text{ MW})\ (0.45) = 608\text{ MW}$$
$$= 34.6 \times 10^6\text{ Btu/min}$$

of energy transfer to the condenser cooling system

Assume: 20°F temperature rise of the cooling water

$$Q = \dot{M}C_p\Delta T$$

$$\dot{M} = \frac{Q}{C_p\Delta T} = 207{,}469 \text{ gpm}$$

This is the circulating or once-through water requirement.

In recirculatory systems:

$$C = \frac{B+D+E}{B+D} = \frac{\text{make-up}}{B+D}$$

where: C = cycles of concentration
B = blowdown
D = drift
E = evaporation

We will calculate the make-up requirement for water systems operating at 2.5, 5.0, and 13.5 cycles of concentration.

From Figure 6.1-3 (RA-352), the following table of make-up requirement for a 1,000 MW power plant versus cycles of concentration can be constructed:

TABLE B.4-1. COOLING SYSTEM MAKE-UP FOR A 1,000 MW PLANT

Cycles of Concentration	Make-up Requirement (10^3gpm)
2	15.6
3	12.2
5	9.8
6	9.2
13	8.2
14	8.2

If it is assumed that at proportional air flow rates the drift and evaporation rates are proportional, then the make-up versus cycles of concentration can be directly scaled from 1,000 MW to a 500 MW plant. Therefore the same general curve would apply. The adjusted valued are shown in Table A.4-2.

TABLE A.4-2. COOLING SYSTEM MAKE-UP FOR A 500 MW PLANT

Cycles of Concentration	Make-up Requirement (10^3gpm)
2	7.8
3	6.1
5	4.9
6	4.6
13	4.1
14	4.1

Plant data from water recycle/reuse studies at the Colstrip (NO-137), Comanche (GA-203), and Montour (CH-387) stations, adjusted by proportionally to equivalent 500 MW plants indicate:

TABLE A.4-3. ADJUSTED PLANT COOLING SYSTEM DATA

	Cycles of Concentration	Make-up Requirement (10^3gpm)
Montour	2.6	8.0
Comanche	5.0	5.5
Colstrip	13.5	4.3

Thus, the two cooling water system data sources show excellent agreement. To fully characterize the plant cooling water systems, it was assumed that the plant evaporation and drift data are characteristic. Table A.4-4 summarizes the four cooling water system characteristics.

TABLE A.4-4. CHARACTERISTIC COOLING SYSTEM OPERATION

System #	Description	Cycles of Concentration	Make-up (gpm)	Blowdown (gpm)	Evaporation (gpm)	Drift (gpm)
1	once-through	1	210,000	210,000		
2	partial recirculatory	2.5	7,000	2,500	4,200	300
3	recirculatory	5.0	5,000	900	3,900	200
4	zero discharge	13.5	4,000	300	3,670	30

A.4-2 Ash Handling System

Assumptions:

1) 75% of the original ash content of the coal forms as fly ash, and 25% forms as bottom ash.

2) 99% of the fly ash is collected in an ESP and wet sluiced to an ash pond as a 5% slurry.

3) Bottom ash is sluiced as one percent slurry.

Table A.4-5 shows the fly ash and bottom ash rates for each of the six example coals.

Assuming 1% bottom ash sluicing and 5% fly ash sluicing, the sluice water requirement is shown in Table A.4-6.

The source for ash sluicing waters varies with power plant water management systems.

- For system #1 it is assumed that raw water is the only source for the ash sluicing water, and that ash sluicing is once-through.

TABLE A.4-5. BOTTOM ASH AND FLY ASH RATES

Coal	Coal Usage Rate lb/hr	Bottom Ash lb/hr	Fly Ash lb/hr
1	646,659	9,700	29,100
2	431,061	6,466	19,398
3	374,342	11,230	33,691
4	395,167	11,855	35,565
5	427,207	5,981	17,943
6	412,348	6,082	18,246

TABLE A.4-6. SLUICE WATER REQUIREMENT

Coal	Bottom Ash Sluice (gpm)	Fly Ash Sluice (gpm)	Total Sluice (gpm)
1	1,939	1,163	3,102
2	1,293	776	2,069
3	2,245	1,347	3,592
4	2,370	1,422	3,792
5	1,196	717	1,913
6	1,216	729	1,945

- For system #2 it is assumed that cooling tower blowdown is the source for the ash sluicing water, and that ash sluicing is once-through. The cooling tower blowdown rate is 2,500 gpm.

Coals #2, #4, and #5 require less ash sluicing water than available through use of cooling tower blowdown. The excess blowdown could be used for other plant requirements or ponded for discharge. No raw water is required for ash sluicing these coals.

Coals #1, 3, and 4 require more sluicing water than available at 2.5 cycles of concentration. In this situation in many operating procedures, cooling tower blowdown would be increased. This reduces the cycles of concentration as shown in the following calculation.

Assuming the same evaporation and drift values, the cycles of concentration to achieve desired blowdown in these cases would be:

TABLE A.4-7

Coal	Desired Blowdown (gpm)	Cycles of Concentration
1	3,102	2.30
3	3,592	2.13
4	3,792	2.08

The raw water requirement is increased to equal the demand for ash sluicing. Passing the water through the cooling tower rather than adding the raw water as make-up directly to the ash sluicing system allows the cooling tower a greater margin of safety in regards to scaling.

A second alternative would be to increase the % solids in the ash slurry. Bottom ash slurries are often in the 1-2% range with many systems operating with 5% solids. Fly Ash Slurries operate in the 5-10% solids range. Often fly ash slurries have higher solids.

If it is assumed that bottom ash is slurried at 2% solids and bottom ash is slurried at 10% solids, the total sluice requirement would be reduced by half. This reduces the required sluice water to:

TABLE A.4-8

Coal	Total Sluice Water Requirement (gpm)
1	1,551
2	1,035
3	1,796
4	1,896
5	957

With the reduced requirement for sluice water, at the desired 2.5 cycles of concentration, the excess blowdown would be routed to a pond before ultimate disposal. Because higher solids concentration slurries lead to handling problems, the power plant could utilize the excess blowdown and operate at as low slurry concentrations as possible. Therefore the ash sluicing requirement is equal to the cooling tower blowdown and therefore there is no additional water make-up requirement for ash sluicing.

A third alternative is that the excess sluice water requirement is met with raw water ash sluice make-up. 1) Because we want to characterize a cooling system operating at 2.5 cycles of concentration (and therefore do not want to reduce cycles of concentration), 2) Because we have defined a 1% slurry of bottom ash and 5% slurry for fly ash as typical operating conditions, and 3) because at least one of the five power plants currently under study by Radian for water recycle/reuse alternatives in an EPA program uses raw water for make-up ash sluice in addition to the cooling tower blowdown, we will choose this third alternative for calculation of water consumption in our study. This choice will not lead to optimal water use (lowering cycles of concentration is a better alternative), but it is a current practice at some power plants and the water requirement will characterize a medium consumption pattern. Table A.4-9 shows the total ash sluice requirement, the raw water ash sluice make-up, and the blowdown for coals #1 through #6 under the third alternative.

TABLE A.4-9. SYSTEM #2 SLUICE MAKE-UP WATER REQUIREMENT

Coal	Total Ash Sluice Requirement (gpm)	Raw Water Make-Up (gpm)	Cooling Tower Blowdown (gpm)
1	3,102	602	2,500
2	2,069	0	2,069
3	3,592	1,092	2,500
4	3,792	1,292	2,500
5	1,913	0	1,913
6	1,945	0	1,945

- For system #3 it is assumed that cooling tower blowdown is the source for ash sluicing water. With this power plant operating at 50 cycles of

- For system #4 it is assumed that cooling tower blowdown is the source for ash sluicing make-up. In this power plant water system the cooling tower operates at 13.5 cycles of concentration, and that the ash sluicing system is total recirculatory. Because the ash settles as a 40-50% solids sludge, and assuming a 5 wt % ash sluice slurry, total recycle translates to 95% recycle of the water stream. Table A.4-11 shows the total sluice water requirement, make-up sluice requirement, and available cooling tower blowdown for each coal.

TABLE A.4-10.

Coal	Total Sluice Water Requirement (gpm)	Make-Up Sluice Water Requirement (gpm)	Available Cooling Tower Blowdown (gpm)	Make-Up Water Requirement (gpm)	Available Gen Serv Blowdown (gpm)	Raw H_2O Req (gpm)
1	3,102	1,551	900	651	563	---
2	2,069	1,035	900	135		
3	3,592	1,796	900	896	563	333
4	3,792	1,896	900	996	563	433
5	1,913	957	900	57		
6	1,945	973	900	73		

Coals #2, #5, and #6 can essentially operate without any raw water make-up requirement by a slight % solids increase. Coals #1, #3, and #4 require make-up water in the quantities shown in Table A.4-10. It is probable that a power plant with a water management program of this nature will probably collect "general service" blowdown (discussed in the following section) and this water will be used as make-up. If raw water is used as make-up, the general services blowdown will probably be fed into the cooling tower circulating stream thus reducing cooling tower raw water demand. We will consider the general services blowdown as ash sluicing make-up. Therefore, coal #1 does not have any additional raw water requirement if a slight increase in solids slurry is made. Coals #3 and #4 require the raw make-up water shown in Table A.4-10.

- For system #4 it is assumed that cooling tower blowdown is the source for ash sluicing make-up. This power plant water system assumes that the cooling tower operates at 13.5 cycles of concentration, and that the ash sluicing system is total recycle. Because the ash settles as a 40-50% solids sludge, and assuming a 5 wt % ash sluice slurry, total recycle translates to 95% recycle of the water stream. Table A.4-11 shows the total sluice water requirement, make-up sluice requirement, and available cooling tower blowdown for each coal.

TABLE A.4-11.

Coal	Total Sluice Water Requirement (gpm)	Make-up Sluice Water Requirement (gpm)	Available Cooling Tower Blowdown (gpm)
1	3,102	155	300
2	2,069	103	300
3	3,592	180	300
4	3,792	190	800
5	1,913	96	300
6	1,945	97	300

It can be seen that no additional make-up requirement is necessary for any of the coals.

A.4.3 General Services Water System

Data from water recycle/reuse studies being conducted by Radian for the EPA were used to characterize this water requirement. Data from the Comanche (GA-203), Montour (CH-387), and Bowen (NO-106) power plants are shown in Table A.4-12.

TABLE A.4-12. PLANT GENERAL SERVICES WATER REQUIREMENT DATA

	Capacity MW	General Services Water (gpm)	gpm/MW
Comanche	700	960	1.37
Montour	1,500	1,900	1.27
Bowen	3,180	11,000	3.46

The Bowen power plant collects the general services water blowdown and feeds it into the cooling tower system. It operates essentially once-through. Therefore, the value for the Bowen general services water will be taken as a high range value. We will assume a value of 1.5 gpm general service water requirement per megawatt rated power plant capacity as a conservative number. For a 500 MW plant this is a raw water requirement of 750 gallons per minute. From Bowen data 1,400 gpm out of 1,900 gpm are collected and routed to a pond (NO-106). This indicates approximately 75% of the general services water requirement might be available for use as cooling tower make-up or ash sluicing make-up. This is 563 gpm in our case . Table A.4-13 shows the general services requirement by system.

TABLE A.4-13.

System	Water Requirement (gpm)	Description
#1	750	once-through
#2	750	once-through
#3	750	recirculated to ash sluicing
#4	187	recirculated to cooling tower

For system #1 - all flows are once-through with no attempt being made to reuse any waters.

For system #2 - we will assume that no attempt is made to reuse this water.

For system #3 - the 563 gpm recoverable general services water is required for ash sluicing make-up. This recycle advantage was considered in reducing the requirement for make-up ash sluicing waters.

For system #4 - The recycle advantage is achieved by combining the recovered general services water with the cooling tower make-up.

A.4.4 Boiler Make-up Water Requirement

Assume that the typical blowdown rate for a drum type steam boiler is 0.1% of the steam generation rate (AY-007). For a 500 MW power plant operating at 37% conversion efficiency, approximately 75 MM Btu/min are used converting water to steam. Assuming a latent heat of vaporization of 1,000 Btu/lb, and a blowdown rate of 0.1% of the steam generation rate, 9.2 gpm of make-up water are required.

A.5 Water Requirements For SO_x Control Strategies

The SO_x control strategies require significant amounts of fresh water for make-up to large circulating streams. The largest make-up requirement is for replacement of water evaporated in the scrubber. Fresh water make-up is also required for loss due to occlusion in solid wastes, prescrubber blowdown, cooling water blowdown, loss with solids drying, loss in product SO_2 streams, and SO_2 conversion process requirements.

A.5.1 Evaporative Scrubber Water Losses

The hot flue gases enter either a prescrubber, or the SO_2 scrubber, and contact a recirculating liquor. The gases are adiabatically cooled and saturated evaporating a portion of the liquor. Fresh water make-up is required to replace this lost water.

Base Case Calculations -

	Mole %	Mol. Wt.	lb/100 moles
CO_2	12.55	44	552.2
H_2O	7.76	18	139.7
N_2	74.55	28	2,087.4
O_2	4.86	32	155.5
SO_2	0.22	64	14.1
NO_x	0.06	--	
			2,948.9

Avg Mol. Wt. = 29.5

flue gas rate = 260,000 scfm per train

$$n = \frac{PV}{RT} = \frac{(14.7 \text{ psia})\ (260{,}000 \text{ scfm})}{(10.73 \ \frac{\text{psia ft}^3}{\text{mole °R}})\ (492\text{°R})}$$

$$= 723.98 \text{ moles/min}$$

$$M_{fg} = 21{,}375.4 \text{ lb/min}$$

Inlet temperature = 310°F

Assume an outlet temperature of 125°F:

$$H_{fg} = M_{fg} C_p \Delta T$$

$$= (21{,}357.4 \text{ lb/min})(0.26 \text{ Btu/°R-lb})(310\text{°R}-125\text{°R})$$

$$= 1{,}027{,}291 \text{ Btu/min}$$

$$H_w = M_w \lambda_w = H_{fg}$$

$$M_w = H_{fg}/\lambda_w$$

$$= (1{,}027{,}291 \text{ Btu/min})/(1{,}020 \text{ Btu/lb})$$

$$= 1{,}007.1 \text{ lb/min of } H_2O \text{ evaporated}$$

$$= 55.95 \text{ moles } H_2O\text{/min}$$

New flue gas composition:

water in original gas = (723.98 moles/min)(0.0776 mol fract H_2O)

$$= 56.18 \text{ moles } H_2O$$

$$\text{Mol \% } H_2O = \frac{55.95 + 56.18}{723.98 + 55.95} \times 100\% = 14.4 \text{ mole \% } H_2O$$

From psychrometric charts @ 125°F, saturated gases contain 12.79 mole % H_2O.

Assume an outlet temperature of 128°F:

$$H_{fg} = M_{fg}\Delta T = M_w/\lambda_w = H_w$$

$$M_w = \frac{MC_p\Delta T}{\lambda w}$$

$$= 55.045 \text{ moles of } H_2O/\text{min}$$

$$\text{new gas composition} = \frac{55.045 + 56.18}{723.98 + 55.045} = 0.14277$$

$$= 14.277 \text{ mole \% } H_2O$$

From psychrometric charts @ 128°F the saturated gas contains 14.105 mole % H_2O.

The agreement is very close. Therefore the outlet temperature is taken as 128°F, and 991 lb/min of water are evaporated per train.

- Model Plant Systems

To calculate the evaporative water losses for each of the model plant systems it is assumed that there is no variation with FGD system. The amount of water evaporated is directly proportional to the amount of flue gas, and therefore, can be scaled directly with plant size. A separate calculation is necessary for each coal type due to different coal usage rates, and flue gas compositions. The evaporative losses for one of four equivalent scrubbing trains for a 500 MW power plant for each of the six coal types is shown in Table A.5-1.

TABLE A.5-1. EVAPORATIVE SCRUBBER WATER LOSSES

	Coal No.					
	1	2	3	4	5	6
T outlet, °F	137	131	128	124	131	132
Water Loss, gpm	130.9	119.1	118.8	119.9	126.5	120.5

A.5.2 Prescrubbing Water Make-up Requirement

In the Wellman-Lord and Magnesia Slurry FGD systems, the flue gases must be scrubbed to remove particulates and chlorides before the gases enter the absorption tower. The evaporative scrubber water loss occurs in the prescrubber, and there is no further evaporation in the absorber. Additional water make-up is required to replace a prescrubber blowdown. The blowdown is taken to maintain desired suspended and dissolved solids concentrations.

For our calculations, it was assumed that 99% of the particulates were removed in an ESP prior to the prescrubber. 100% of the remaining particulates, and 100% of the chlorides are assumed to be removed from the flue gases in the prescrubber. The particulates are sluiced in a 5% solids slurry to the ash pond, and the limiting concentration of dissolved solids is 20,000 ppm.

Assumptions:	make-up water	40 ppm chlorine
	flue gases	30 ppm chlorine

- Dissolved Solids Blowdown Requirement

Chlorine Balance

chlorine input = chlorine output

$$F \cdot f_{cl} + M \cdot M_{cl} = B \cdot b_{cl} \qquad (1)$$

Water Balance

Water in = Water out

$$M + \cdot F\ f_{H_2O} = B + E + F \cdot f_{H_2O}$$

$$M = B + E \qquad (2)$$

M = make-up, lb/min

B = blowdown, lb/min

E = evaporation, lb/min

F = flue gas rate, lb/min

f_{cl} = weight fraction of chlorine in the flue gas

f_{H_2O} = weight fraction of water in the flue gas

b_{cl} = weight fraction of chlorine in the blowdown

M_{cl} = weight fraction of chlorine in the make-up water

Combining (1) and (2) and solving for B:

$$B = \frac{E \cdot M_{cl} + F \cdot f_{cl}}{b_{cl}} \qquad (3)$$

Table A.5-2 shows the results of the calculations for each of the coals.

TABLE A.5-2. DISSOLVED SOLIDS BLOWDOWN REQUIREMENT

Coal	F lb/min	E lb/min	B lb/min
1	24,747	4,364	45.85
2	21,768	3,972	40.60
3	21,357	3,964	39.97
4	21,088	4,000	39.63
5	23,119	4,220	43.12
6	22,146	4,020	41.26

f_{cl} = 30 ppm = 3×10^{-5} lbd/lbF

M_{cl} = 40 ppm = 4×10^{-5} lbd/lbM

B_{cl} = 20,000 ppm = 0.02 lbd/lbB

- Suspended Solids Blowdown Requirement

Table A.5-3 shows the water requirement for sluicing the particulates removed in the prescrubber. The sluice is 5% solids.

TABLE A.5-3. SLUICE WATER REQUIREMENT

Coal	Water Requirement, lb/min
1	97
2	65
3	112
4	119
5	60
6	61

Thus, the prescrubber blowdown is limited by the desired suspended solids concentration. For the base case, for all four scrubbing trains, the make-up requirement is 54 gpm.

A.5.3 Water Loss Due to Occlusion in the Solid Waste

The lime, limestone, and double alkali FGD systems produce a by-product solid waste. This solid waste is assumed in this study to be settled or dewatered to a 60% solids sludge. Water loss due to occlusion in the sludge requires fresh water make-up.

- Base Case Calculation

$$\text{gas rate: } 260{,}000 \text{ scfm} \times \frac{1 \text{ lb-mole}}{359 \text{ scf}} = 723 \text{ lb mole/min}$$

SO_2 content: 0.0022 volume fraction

Assume 90% SO_2 removal:

$$(0.90)\left(723 \frac{\text{lb moles}}{\text{min}}\right)(.0022) = 1.43 \frac{\text{mole } SO_2}{\text{min}}$$

Assume 25% oxidation and coprecipitation of sulfite and sulfate:

$$\begin{array}{l} 0.36 \text{ moles } CaSO_4 \\ 1.07 \text{ moles } CaSO_3 \end{array} \qquad \begin{array}{r|l} 1.07 & .15 \text{ mole } CaSO_4 \\ \hline & .85 \text{ mole } CaSO_3 \end{array} = 0.19 \text{ moles}$$

$$\therefore \; 1.26 \text{ moles } Ca(.85SO_3 \cdot .15SO_x) \cdot \tfrac{1}{2}H_2O$$

$$0.17 \text{ moles } CaSO_4 \cdot 2H_2O$$

Assume a lime stoichiometry of 1.05:

excess lime regenerant, recarbonated to $CaSO_3$

$CaCO_3 = (0.5)(1.43) = 0.7$ lb moles $CaCO_3$

Sludge solids:

$Ca(.85\ SO_3 \cdot .15\ SO_4) \cdot \frac{1}{2}H_2O$	(1.26)(130) = 163.8 lb/min
$CaSO_4$	(0.17)(172) = 29.2 lb/min
	(0.07)(100) = 7.0 lb/min
	200.0 lb/min

these solids contain 17.4 lb/min of water of hydration

Assume 60% solids sludge:

occluded water = 133 lb/min + 17.4 lb/min

= 150.4 lb/min

- Model Plant Systems Calculations

Each model plant system, excluding variations in power plant size, requires a separate calculation. Because the amount of water occuluded in the solid waste is directly proportional to the power plant size, these systems' water losses can be scaled from the values calculated for 500 MW plants. The results of the calculations are shown in Table A.5-4.

Cooling Water Blowdown

The Wellman-Lord condenser cooling system, and the product acid cooling system in the contact sulfuric acid plant,

TABLE A.5-4. WATER LOSS: OCCLUDED WITH SOLID WASTE

Case	%S Removal	FGD System	Coal	Water Loss (lb/min)
1 a	76.4	L	#3	128
	76.4	LS	#3	141
	76.4	DL	#3	128
	76.4	DS	#3	141
b	88.2	L	#4	303
	88.2	LS	#4	332
c	not applicable			
d	not applicable			
e	62.5	L	#5	46
	62.5	LS	#5	51
2 a	90	L	#3	150
	90	LS	#3	165
	90	DL	#3	150
	90	DS	#3	165
	90	L	#4	309
	90	LS	#4	339
	90	DL	#4	309
	90	DS	#4	339
b	90	L	#1	49
	90	LS	#1	53
	90	L	#2	35
	90	LS	#2	38
3 a	70	L	#2	27
	70	LS	#2	30
	70	L	#1	38
	70	LS	#1	42

(Continued)

TABLE A.5-4. WATER LOSS: OCCLUDED WITH SOLID WASTE (Continued)

Case	%S Removal	FGD System	Coal	Water Loss (lb/min)
b_1	85	L	#5	98
	85	LS	#5	110
b_2	90.8	L	#6	196
	90.8	LS	#6	214

L - LIME

LS - LIMESTONE

DL - DOUBLE-ALKALI: lime regenerant

DS - DOUBLE-ALKALI: limestone regenerant used for conversion of product SO_2 streams, require fresh water make-up due to losses of evaporation, drift, and blowdown. These values were scaled from the requirements presented by McGlammery, et al (MC-147).

A.5.5 Loss with Solids Drying

The Wellman-Lord and Magnesia Slurry FGD systems contain solids drying operations. The water loss associated with these operations is relatively small, but requires make-up to the system. This make-up requirement was calculated from water balances (Section B.6) and 100% loss was assumed. The results were scaled for each model plant system.

A.5.6 Loss in the Product SO_2 Stream

The product stream from the Wellman-Lord process was assumed to contain 10% water. The quantity of water lost was calculated by sulfur removal requirement, and scaled for each of the model plant systems.

A.6.0 WATER BALANCES FOR FGD SYSTEMS

Water balances were calculated for the base case conditions for the five FGD systems studied in this project. Tables showing the flows, and water compositions of the various streams are given in Section 5 of the report, when discussing each FGD water system. The method of calculation is straightforward, and the major assumptions are listed below the table for each system.